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16. ABSTRACT The proceedings document the eighth in a series of symposiums (the second consponsored by EPA and the Electric Power Research Institute) on flue gas desulfurization (FGD). It provided a forum for the exchange of technical developments and regulatory information on FGD systems and processes applicable to utility and industrial boilers. The keynote address described the National Acid Precipitation Assessment Program. Other opening-session presentations were on the status of air quality legislation/regulations, the status of solid waste regulations, and trends in commercial application of FGD technology. Subsequent sessions dealt with FGD economics, materials of construction, dry furnace absorbent injection, dual alkali FGD, flue gas treatment (combined SOx/NOx removal), FGD chemistry, the architect-engineer's role in FGD applications (panel discussion), limestone/organic acid, waste disposal/utilization, and dry FGD systems. Participants represented electric utilities, equipment and process suppliers, state environmental agencies, coal and petroleum suppliers, EPA and other federal agencies, and research organizations.

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Flue Gases	Calcium Carbonates	Stationary Sources	21B	
Desulfurization	Organic Acids	Acid Precipitation	07A, 07D	07C
Waste Treatment				
Economics			05C	
Sulfur Oxides			07B	

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PREFACE

These proceedings for the Eighth Symposium on Flue Gas Desulfurization constitute the final report submitted to the Industrial Environmental Research Laboratory, U.S. Environmental Protection Agency, Research Triangle Park (IERL-RTP), North Carolina, and to the Coal Combustion Systems Division, EPRI, Palo Alto, California. The symposium was conducted in New Orleans, Louisiana, November 1 to 4, 1983.

This symposium was designed to provide a forum for the exchange of technical and regulatory information and developments on the systems and processes applicable to utility and industrial boilers. In the opening session, a keynote address was given on the status and outlook for the national acid precipitation assessment program. Presentations were also made on the status of air quality legislation and regulations, the current and projected regulations of the resource conservation and recovery act, and trends in commercial application of flue gas desulfurization (FGD) technology. There were subsequent technical sessions on economics, materials of construction, dry furnace absorbent injection, dual-alkali, flue gas treatment (combined SO_x/NO_x) removal, the role of the architect-engineer as the middleman between the utility and FGD supplier (discussion), FGD chemistry, limestone/organic acid, waste disposal/utilization, and dry FGD systems, pilot plant test results and full scale installations. Participants represented electric utilities, equipment and process suppliers, state environmental agencies, coal and petroleum suppliers, EPA and other Federal agencies, and research organizations.

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Chairman: J. David Mobley
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PROCESS TROUBLESHOOTING AT A UTILITY LIMESTONE FGD SYSTEM

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ABSTRACT

Central Illinois Light Company's Duck Creek Unit No. 1 experienced significant reliability and operating problems with its limestone FGD system following start-up in 1979. CILCo entered into a testing and evaluation program co-funded by the Electric Power Research Institute in late 1981 to verify the feasibility of using additives to improve the system's low SO_2 removal efficiency. A second objective of the work was to improve the system's reliability, which averaged slightly better than 60 percent prior to testing. Severe mist eliminator scaling that was causing routine scrubber outages was primarily responsible for the reliability problems. The eight month program that followed involved extensive chemical process troubleshooting.

Two types of organic acid buffers were tested along with the addition of magnesium oxide during the period. Both dibasic acid (DBA) and the magnesium proved capable of enhancing SO_2 removal to levels sufficient to maintain the unit in compliance with the 1971 SO_2 NSPS. An economic analysis was performed based on the data collected during this testing. The cost study compared capital and annual operating and maintenance costs for each option over the remaining life of the plant.

Work was conducted concurrent with the additive testing to solve the mist eliminator scaling. The cause of the scaling was identified and effectively stopped by switching to a fresh water wash, adjusting the wash sequence, and improving limestone utilization. Improvements in limestone utilization were accomplished by optimizing the operation of the mill circuit to provide a finer, more reactive product. Since completion of the program, the FGD system has consistently achieved reliability numbers greater than 95 percent while at the same time lowering operating and maintenance costs.

INTRODUCTION

Duck Creek #1 is a bituminous coal fired unit which is owned and operated by Central Illinois Light Company (CILCo). The boiler is rated at 400 megawatts and began commercial operation in 1976. It fires a high sulfur Illinois coal and is subject to the 1971 utility NSPS which limits the units SO_2 emissions to 1.2 pounds per million Btu boiler heat input. The FGD system consists of four Riley Environeering rod deck scrubber modules each designed to treat 25 percent of the maximum gas flow.

Since startup of its scrubbers, CILCo has experienced many of the same operating problems seen at other utility limestone FGD systems. The primary difficulties have been low SO_2 removal, chemical scaling in the mist eliminators, materials and liner failure, and low reagent utilization. By 1981, the utility was facing regulatory pressure because of SO_2 compliance violations and was reporting FGD reliabilities of only slightly more than 60 percent. In addition, the low reagent utilization was leading to accelerated filling of the FGD waste disposal pond.

CILCo contracted Radian Corporation in late 1981 to conduct a test program whose primary objective was to demonstrate the feasibility of using an organic buffer to improve the FGD system SO_2 removal efficiency. Severe weather hampered the testing but the initial results were encouraging. Further test work was delayed until April following a month long plant outage. As the testing progressed after startup in April, the scope of work expanded to include investigating the mist eliminator scaling problem and ways to improve limestone reagent utilization. The Electric Power Research Institute became involved at this stage and funded the cost of the remainder of the program. This included conducting a test to determine the feasibility of using magnesium as an alternative to the organic buffers and completing a comparison of the economics of the available options for improving SO_2 removal efficiency. The results of the test work and cost study are discussed in this paper along with a description of the FGD system. Finally, observations are presented from a recent Radian sampling trip to document the systems performance a year after the end of the Radian/EPRI program.

PROGRAM OBJECTIVES AND SCHEDULE

The overall objectives of the test program were to improve both the SO_2 removal efficiency and reliability of the Duck Creek FGD system. The organic buffer and magnesium testing were intended to demonstrate effective means of increasing the removal efficiency of the scrubbers. This work gathered process and chemical data over a variety of system operating conditions and proceeded in three phases. The first phase involved gathering baseline data over a one week period to characterize the performance of the scrubbers before additive testing began. The information was collected to quantify the magnitude of the SO_2 removal problem and to serve as a point of reference for data collected later.

The organic acid demonstration test work which made up phase two began in early May, 1982 and continued for four months until the end of the testing in

August. Phase three, the magnesium testing, began the first of August and concluded four weeks later.

To accomplish the second objective to increase the reliability of the FGD system, work to solve the severe mist eliminator scaling was undertaken in early May. This proceeded concurrent with the additive testing and lasted throughout the 1982 test phase. Improving limestone reagent utilization was important in solving the mist eliminator scaling problem. This work also ran concurrent with the additive testing and lasted through the program.

SYSTEM DESCRIPTION

The limestone FGD system at Duck Creek was designed to treat the entire volume of flue gas from the boiler and limit SO_2 emissions to below 1.2 lbs per MMBtu. At full load operation, approximately 400 MW of power are being generated resulting in a total flue gas flow of 1.4 million acfm at scrubber exit conditions (130°F, saturated). The boilers induced draft fans discharge into a common plenum to which the inlet ducts of the four modules are connected. Distribution of the flue gas is determined by the pressure drop across each module and is not regulated by dampers. The flow to each may be adjusted by manually removing rods from or adding rods to any of the seven decks inside each tower. The SO_2 concentration of the 300°F gas as it enters the towers averages about 2500 ppm. A simplified flow diagram showing one of the four modules and the reagent preparation plant and waste disposal is shown in Figure 1. Reagent preparation and waste disposal are common to the FGD system.

Limestone preparation is accomplished with a 10' x 18' wet ball mill which was originally designed to grind 40 tons per hour of stone. Four 6 inch diameter hydroclones are used to control the product size distribution. Columbia Quarry supplies the stone which is mined underground from the Kimmswick formation. Product slurry at about 25 weight percent solids is stored in a 125,000 gallon tank prior to being fed to the scrubber reaction tanks.

Limestone feed to each reaction tank is controlled based on a pH feedback loop. The pH setpoint in the reaction tanks can be maintained within ±0.05 units with control of the limestone slurry in an on/off mode. The 150,000 gallon working volume of the reaction tanks provides the necessary time for precipitation of the sulfur salts and dissolution of the limestone. Reaction tank slurry is pumped to the top of the contactors at a rate of about 15,000 gpm where it passes countercurrent to the flue gas. At full load conditions, this results in a liquid-to-gas ratio of slightly more than 40 gals/1000 acf. Gas-liquid contacting to promote SO_2 mass transfer is achieved with the seven rod decks. Gas side pressure drop across the scrubbers is approximately 12 inches H_2O at design flow rates.

A two-stage horizontal mist eliminator (constructed of Hastelloy G) is located downstream of the last rod deck in each absorber to remove entrained slurry. Periodic washing of the mist eliminator removes collected solids. The majority of the wash water is routed to the disposal pond with a small amount entering the recycle loop. Flue gas exits the system without reheat through a lined wet stack.

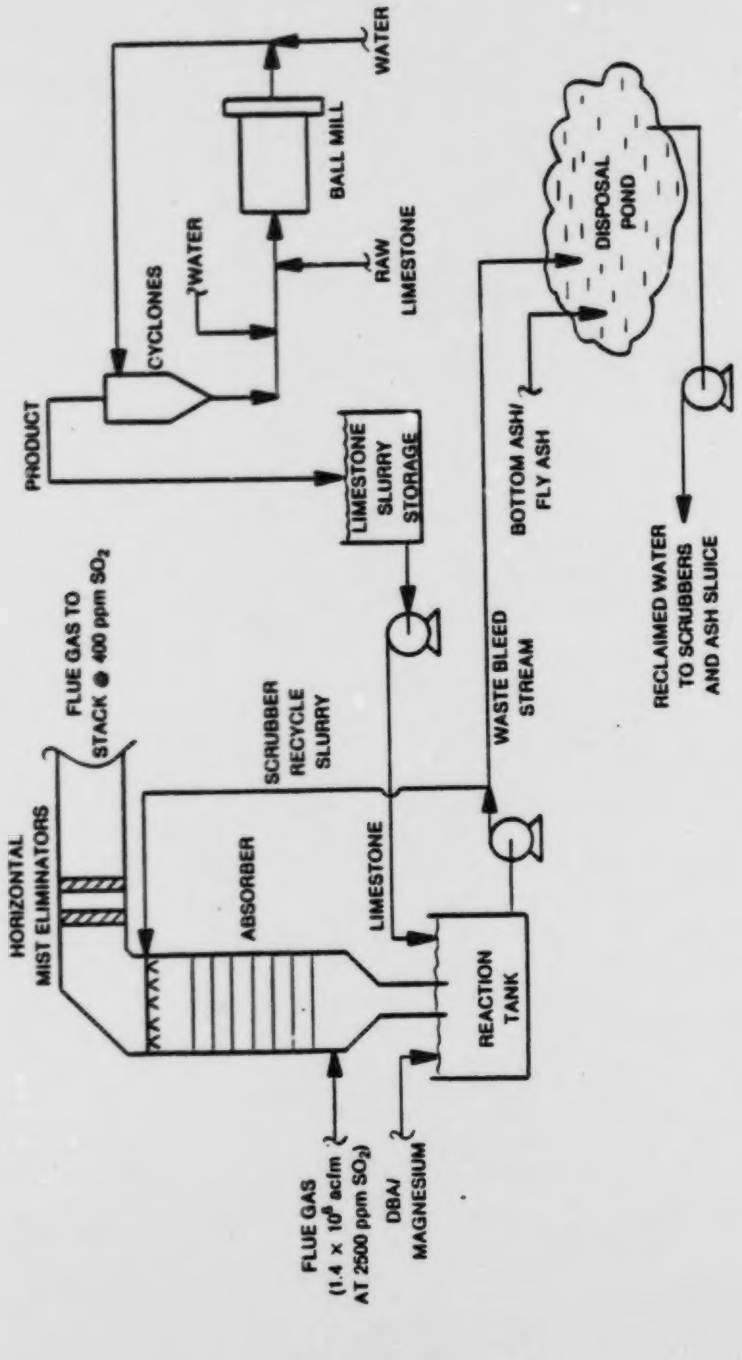

Figure 1. Duck Creek FGD System

Solids content in the reaction tank is controlled at 12 weight percent by adding dilution water reclaimed from the disposal pond. Waste slurry is pumped from each reaction tank to the pond. This pond also serves as disposal for the boiler bottom and fly ash as well as a final repository for all other plant wastes.

DISCUSSION OF RESULTS

This section presents the key results of the four month period of testing in 1982 and the 1983 evaluation trip. The results of the additive testing are presented first in the SO_2 Removal subsection. This is followed by results of the system reliability improvement work which are presented in the Mist Eliminator Scaling and the Reagent Utilization subsections. The Cost Study which investigated the economics of the alternative additives is discussed next. Finally, observations made one year following the 1982 testing are presented in the 1983 Evaluation Trip subsection.

SO_2 REMOVAL

Baseline Testing

A characterization of the FGD system was completed at the beginning of the program to document the operation and performance of the FGD system with particular emphasis on SO_2 removal. The testing was done prior to the additive testing to define the baseline characteristics of the scrubbers. Data were collected at various reaction tank pH setpoints and over a range of unit loads. The results are plotted so that the effect of pH and gas flow on the scrubbers SO_2 removal can be evaluated. Figure 2 presents the relationship observed between reaction tank pH and the SO_2 removal efficiency of the scrubbers. There is a strong correlation between pH and removal up to about 93 percent removal indicating that liquid phase alkalinity is limiting the scrubbers removal efficiency. The data shown in the figure were taken with the unit operating at or above 90 percent of its maximum to minimize the confounding effect of gas flow on the results. The figure shows that controlling at a reaction tank pH of 5.9 results in a removal efficiency of roughly 85 percent. This removal efficiency is the approximate level the scrubbers are required to operate at to achieve the compliance emission limit of 1.2 lbs SO_2 per million Btu. As will be discussed in the next two subsections, this high of a pH leads to scaling in the mist eliminators (ME) and poor reagent utilization.

Figure 3 shows the effect of unit load (actually gas flow rate) on the stack SO_2 emission rate. Above a load of about 320 MW, the emission rate begins to exceed the compliance limit. At a load of 200 MW, the average SO_2 removal efficiency approaches 90 percent. At full load (400 MW), the removal drops off to about 75 percent. The data were taken at an average pH setpoint of 5.7.

Organic Acid Testing

Pilot scale testing at the EPA-IERL facility, prototype testing at the Shawnee Alkali FGD Test Facility, and the full scale demonstration program at

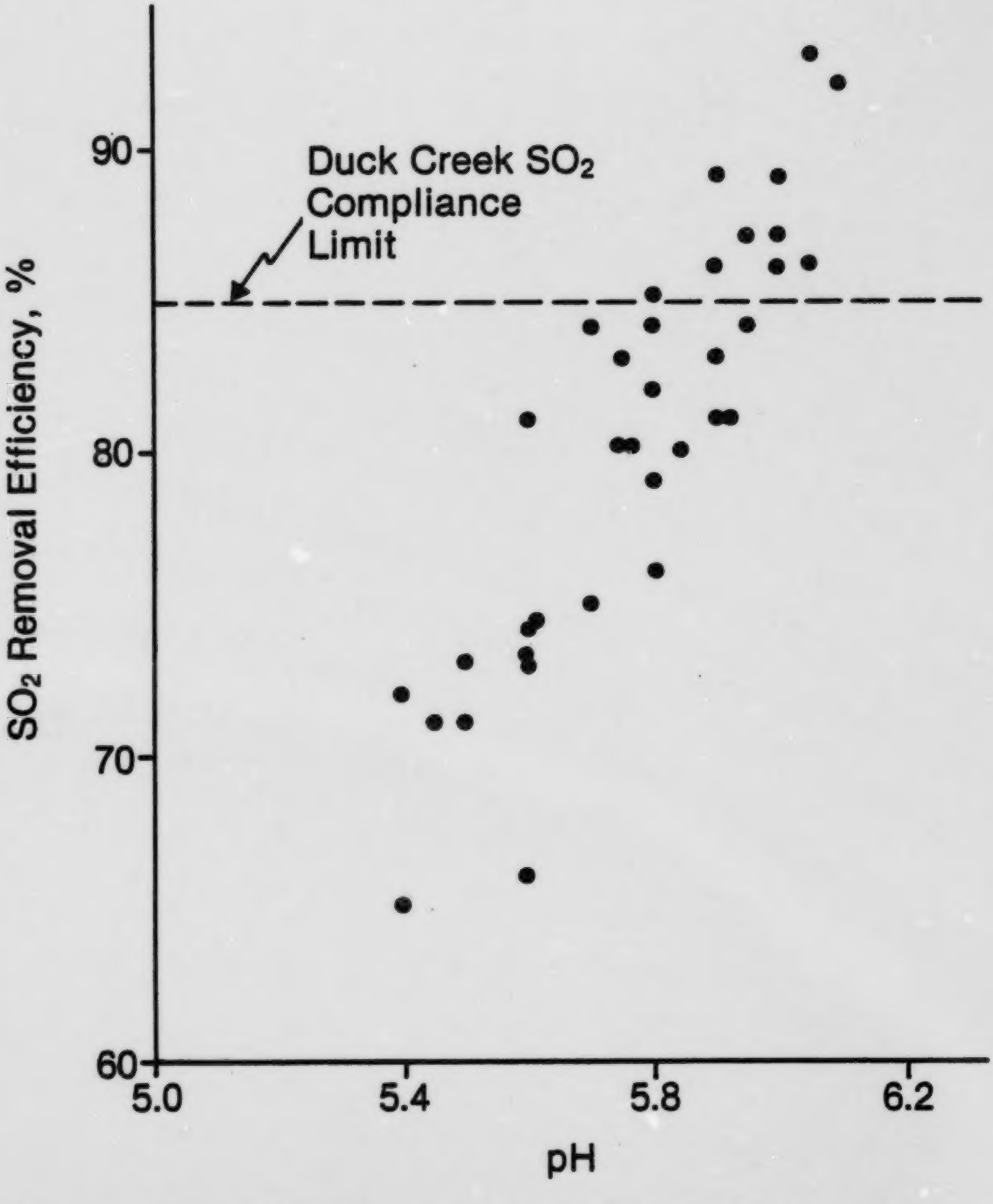

Figure 2. SO_2 removal as a function of scrubber pH - baseline test period

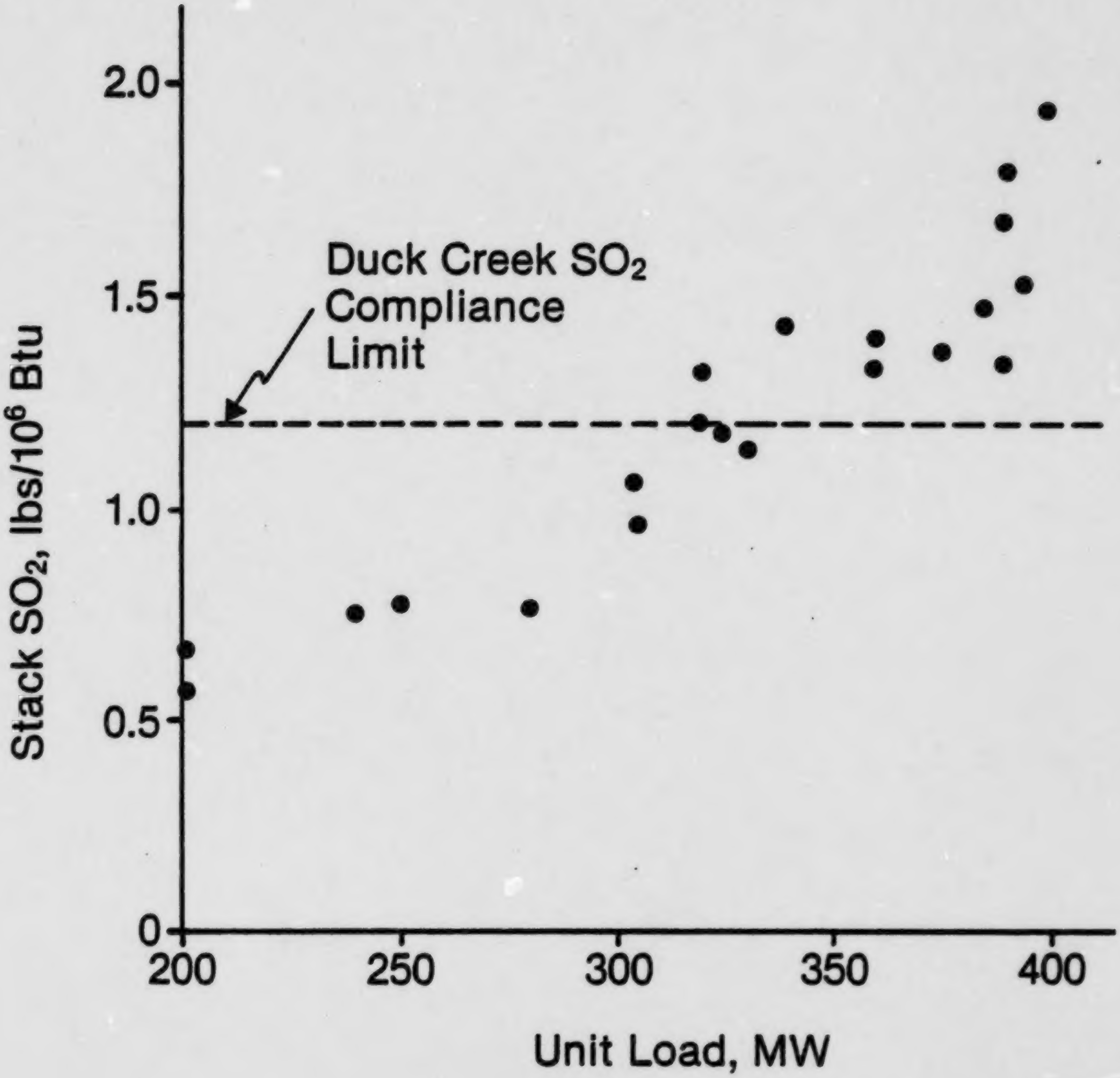

Figure 3. Unit load vs. stack SO_2 emissions - baseline test period

City Utilities Southwest Power Plant documented the feasibility of using organic acids to enhance the SO_2 removal efficiency of lime/limestone scrubbers. The organic acids act as buffers when dissolved in the liquid phase of the reaction tank slurry. As buffers, they prevent a large drop in pH at the gas-liquid interface which naturally occurs due to the absorption of SO_2. This improves the mass transfer rate of SO_2 from the gas to the liquid by reducing the liquid phase resistance. The higher the dissolved concentration of the organic acids, the greater the improvement in mass transfer and therefore the better the SO_2 removal efficiency. The upper limit in improvement is set by the gas-liquid contacting efficiency which determines the magnitude of the gas phase resistance. Of course, the higher the dissolved buffer concentration, the higher the addition rates and therefore annual costs of the raw material. Testing was conducted to determine the concentration of additive required to achieve the desired SO_2 removal efficiency.

To accomplish this, the organic buffers tested were pumped directly into two of the four reaction tanks at measured rates. The concentration in the liquid phase in each was measured by an acid-base titration procedure. At the same time, SO_2 removal, unit load, and pH data were collected for correlation purposes.

Two types of buffers were tested. The majority of the testing was with a material commonly referred to as dibasic acid or DBA, which is a coproduct resulting from the manufacture of adipic acid. It is an aqueous mixture of adipic, glutaric, and succinic acids, all equally as effective a buffer in a limestone scrubber. The other type of organic additive tested was composed primarily of adipic acid and hydroxycaproic acid, a monobasic molecule. Testing with this material, commonly called acid water, was limited due to severe reaction tank foaming caused by a component in the mixture.

Figures 4 and 5 present the observed effect of DBA concentration on SO_2 removal for Modules A and B, respectively. All data shown were taken at unit loads above 360 MW (90% of maximum). The data taken while the reaction tanks were being controlled at a pH of 5.4 to 5.7 is separated from that taken during operation in the 5.7 to 6.0 range. For operation in the higher pH range, Module A required approximately 200 ppm DBA to maintain removal efficiencies above 85 percent. Less DBA was needed in Module B (100 ppm) to get 85 percent removal in the higher pH range. Data were collected at the lower pH's for estimating the trade-off in limestone versus DBA consumption. For operation in the lower pH range, approximately 400 ppm DBA is required for Module A and about 300 ppm for Module B for 85 percent removal. The reason for the better performance for B is unknown since liquid recycle rate and gas flow measurements for each were virtually identical.

Because Duck Creek uses a disposal pond instead of a thickener/filtration step for dewatering waste scrubber sludge, little of the buffer added is returned to the scrubbers. The pond has roughly 100 times the volume of a typical thickener. Estimates of the loss rate of DBA from the reaction tanks in the waste slurry were compared with the actual feed rates to determine the extent of DBA losses through chemical degradation, coprecipitation with the sulfur salts, or in fugitive streams. The results showed that

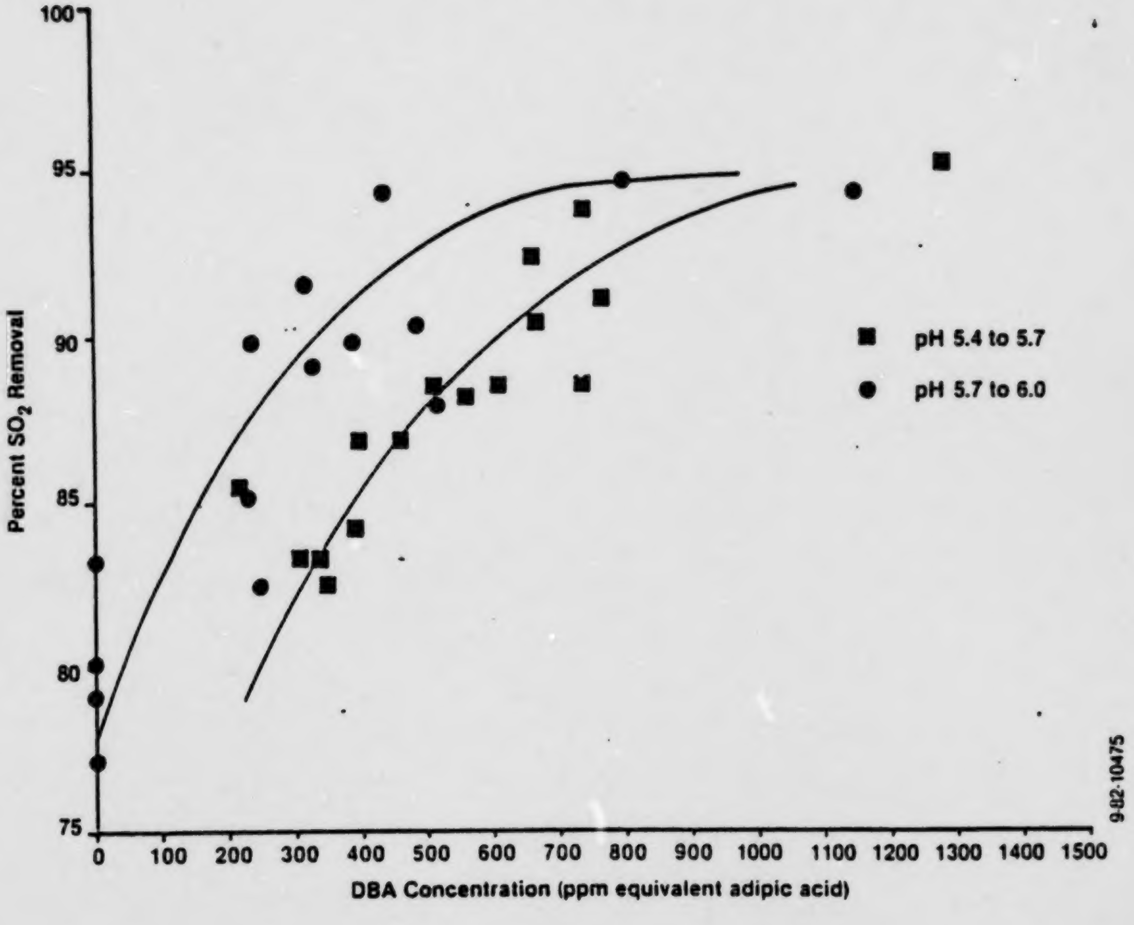

Figure 4. Effect of DBA concentration on percent SO_2 removal--Module A

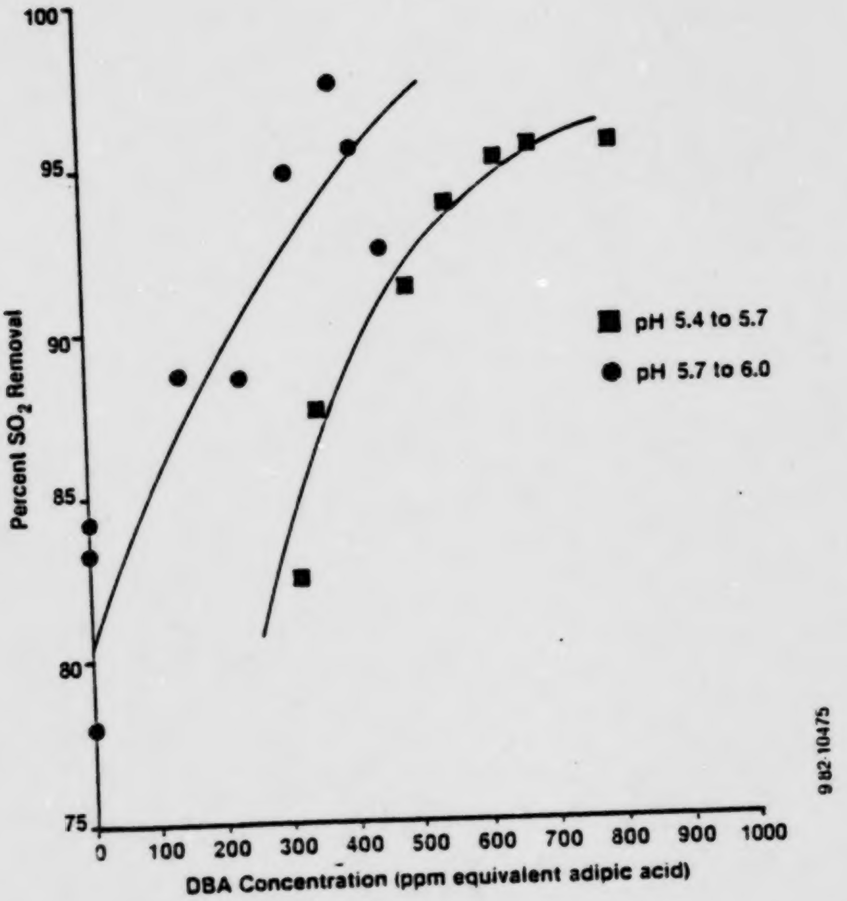

Figure 5. Effect of DBA concentration on percent SO_2 removal--Module B

within the accuracy of the measurements all of the material fed could be accounted for in the waste slurry which left the reaction tank. This information is important because for closed loop FGD systems (those designed with thickener/filtration solids dewatering) loss rates over 5 times higher than filter cake liquor loss rates have been observed for nearly equivalent DBA concentrations. This difference is probably due to the relatively short liquid phase residence time in the CILCo system (10 hours) compared to a closed loop system (weeks).

Based on these results, annual DBA consumption rates can be estimated for the Duck Creek FGD system. Required concentrations for the two modules not tested (C and D) are assumed to be the same as for Module A. Assuming 300 days per year online and 12 hours per day operation of the unit at or above 90 percent of full load during which time DBA must be fed, 1.26×10^6 pounds of DBA (dry basis) are required for a low pH setpoint (5.5) and 630,000 pounds for a high pH setpoint (5.8). Using a DBA delivered cost of 29.5 cents per pound, dry basis, the annual cost difference between the two is roughly $200,000. The section on Limestone Utilization discusses the tradeoffs with annual limestone consumption.

Magnesium Testing

Testing to examine the effectiveness of using magnesium to enhance scrubber SO_2 removal efficiency was conducted in August, 1982. Dolomitic lime was used as the source of magnesium to determine if it was a cost effective alternative to magnesium hydroxide. A detention slaker was installed to hydrate the lime (calcium fraction only) prior to its being fed to the scrubber reaction tank tested. In the reaction tank the magnesium oxide dissolves, increasing the slurry liquid phase magnesium concentration. Because magnesium sulfite is much more soluble than calcium sulfite, the liquid phase sulfite concentration increases depending upon how much magnesium is present. At the same time, the magnesium also pairs with the dissolved chloride and sulfate ions which reduces its effectiveness. The increase in sulfite improves the liquid phase alkalinity of the slurry by increasing the $SO_3^=$ species concentration and the $MgSO_3$ complex concentration. Both act as buffers in the same way as the organic acids, promoting the mass transfer of SO_2 from the gas to liquid phase.

Figure 6 presents the relationship observed between SO_2 removal and dissolved magnesium concentration. No noticeable improvement in SO_2 removal was seen below about 1000 ppm of dissolved magnesium. This is probably due to the chloride and sulfate association. Approximately 1500 ppm of dissolved magnesium was sufficient to maintain Module A at 85 percent SO_2 removal. Mass balance calculations verified that all of the solid magnesium added to the reaction tank dissolved. The dolomitic lime appears to be a viable means of adding magnesium to the scrubbers, therefore. During the testing, the reaction tank pH was controlled at a relatively high pH (5.8 to 6.0) to maximize the $SO_3^=$ and $MgSO_3$ complex species concentrations. Dropping the reaction tank pH to 5.4 would theoretically shift the sulfite-bisulfite equilibrium such that about a 30 percent reduction in concentration of the $SO_3^=$ and $MgSO_3$ species would result. Actual test data were not available to verify this because of difficulties in operating the slaker for

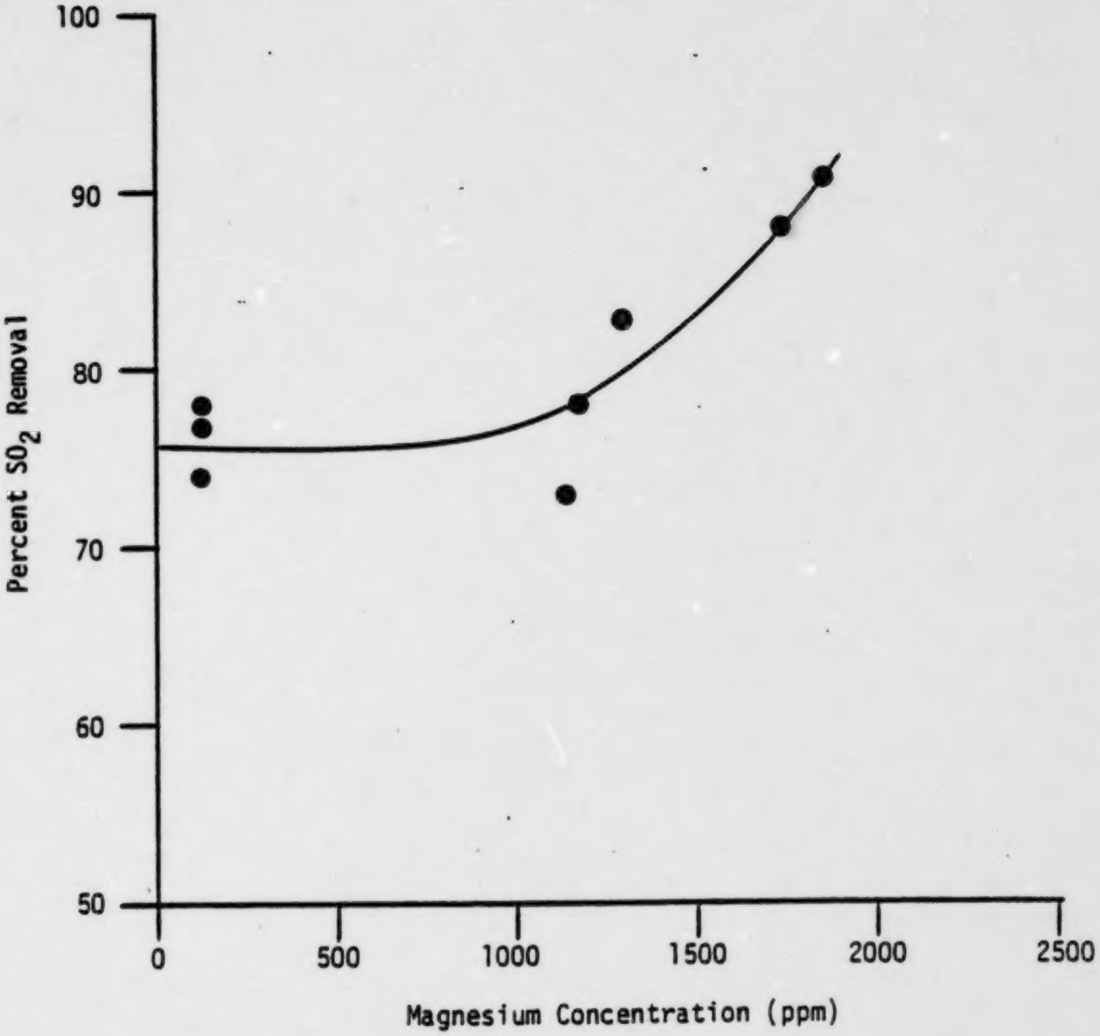

Figure 6. Effect of magnesium concentration on percent SO_2 removal--Module A

extended periods. However, if accurate, a 30 percent increase in the magnesium feed rate would be required to offset the reduction. This would be in addition to the increase in magnesium required to make up for the loss in scrubbing efficiency due to the lower pH operation. Recall that roughly a two fold increase in DBA was required in dropping from a pH setpoint of 5.8 down to 5.5. The reason for this drop in efficiency is a combination of the reduction in liquid phase alkalinity and of the reduction in solid phase alkalinity (limestone) due to less excess limestone present in the slurry at lower pH's. The tradeoff of low pH operation/improved limestone utilization is not favorable for magnesium as it is for DBA.

Mass balance calculations also showed that due to the method of solids disposal at Duck Creek, high feed rates of magnesium would be required until steady state could be reached in the entire system. Depending upon the exact volume of the disposal pond, this could require a number of years. To improve the economics of magnesium-enhanced scrubbing, recycle of scrubber blowdown liquor was investigated. The ability of hydroclone classifiers to partially dewater the scrubber wastes was tested. Because of the extremely fine size of the solids, small diameter classifiers were necessary to make the desired separation. Results showed that the slurry could be concentrated from 10% solids up to about 40% solids, reducing the magnesium consumption by a factor of 4. The effect of a recycle step on the operation of the scrubbers is unknown and could not be investigated short of major mechanical modifications. The reduction in DBA consumption would be somewhat less than that for magnesium since the fraction of organic buffer chemically degrading or coprecipitating with the sulfur would expect to increase as its residence time in the absorber loop increased. The effect of this recycle step on the economics of these two alternatives is discussed in the Cost Section later.

MIST ELIMINATOR SCALING

Prior to the test program, mist eliminator (ME) scaling was causing a severe reliability problem for the Duck Creek FGD system. In an attempt to control the scale, each module's mist eliminator was cleaned weekly by a crew of maintenance men. Usually the scale growth rate was so fast that within a few days, solids pluggage would cause a significant increase in the ME gas pressure drop. The restriction resulted in a maldistribution of gas flows between the four absorbers. Higher gas flows through a recently cleaned tower would result because of the lower resistance to flow through its ME. The higher gas flows would reduce that absorbers SO_2 removal efficiency as well as carry up greater quantities of mist to the ME. The increase in removal efficiency due to the lower gas rates through the restricted towers did not completely offset this reduction and overall the removal would drop. Trouble with completely isolating the flue gas from the inside of the towers during cleaning periods required that some gas bypass the system. Therefore, the mist eliminator scaling was not only causing reliability problems, but it was also indirectly causing compliance problems.

A brief period of time was spent early in the program testing the effectiveness of adding a gypsum scale inhibitor to the ME wash water. Even at higher than recommended dosage rates, no effect on either the formation rate or strength of the scale were seen.

At the same time, work began to identify the causes of the scaling and potential solutions for eliminating it. Regular inspections of the mist eliminators verified the rapid buildup of solids on their surface. Analysis of the solids showed that they were primarily gypsum with small percentages of calcium carbonate and calcium sulfite present. The wash rates were estimated to be about 1100 pgm with the front and back of the first stage ME and front side only of the second state being washed. With the ME's having a front face area of about 480 ft^2, this resulted in a specific wash rate of about 0.75 gpm/ft^2. Each mist eliminator was washed for 5 minutes every twenty minutes.

During the plant outage prior to the test program each ME was thoroughly cleaned and all wash nozzles and headers were cleaned or replaced. Several weeks into testing, a significant number of spray nozzles were requiring replacement each week. The nozzles were plugging with scale themselves. Inspection of the inside of the spray headers showed that a thick gypsum scale was growing on the walls. The pluggage was reducing wash rates and this combined with the plugged nozzles was accelerating scale formation on the ME's. The wash water which was reclaimed disposal pond liquor was analyzed for major cations and anions and the results input to the Radian Inorganic Equilibrium Program. The model predicted a gypsum relative saturation of 1.9 which is significantly higher than the level at which gypsum scaling usually initiates (1.4 relative saturation). From this analysis, it was obvious that the reclaimed pond liquor could no longer be used for washing the ME's.

The only other source of water available was fresh water from the cooling lake. Using it at a rate of 1100 gpm would cause significant overflow of the disposal pond which was not allowed. Results of previous work with mist eliminators at the Shawnee Test Facility indicated that minimum ME wash rates could be achieved using a combination of fresh water (or a water low in TSS, dissolved calcium, and dissolved sulfate), good limestone utilization, and sufficient wash intensity. Based on the recommended wash rates and durations and on the dimensions of the Duck Creek ME's, a total of only 100 gpm of fresh water would be needed to keep them scale free if the limestone utilization was kept above 85 percent.

After the necessary piping changes were made to switch to fresh water, durations and frequencies were slowly reduced to find the minimum required to keep the mist eliminators clean. To help offset the impact on the plant water balance, scrubber recycle pump seal water was closely monitored to eliminate excess flush rates and hoses used for general cleaning around the plant were switched from fresh water to pond reclaim water. The wash rate had been reduced to about 300 gpm by the end of August. Scaling and pluggage of the mist eliminators were virtually eliminated.

Further optimization of the washing was possible, though. The front of the first mist eliminator was not being washed with the recommended intensity. More extensive piping and valve changes were required for which there was not sufficient time during the 1982 testing. The Shawnee results called for six minutes of washing every four hours at a specific wash rate of 1.5 gpm per ft^2 of ME face area for the front of the first ME (nozzle pressure 41 psig). The recommendation for the back of the front ME was 4 minutes of wash every 8 hours at a specific rate of 0.55 gpm/ft^2 (nozzle pressure 13 psig). There was no recommendation for the second bank ME wash, but it was assumed to be equal to or less than that for the back of the first bank. The piping configuration for the Duck Creek fresh water wash resulted in the front and back of the first ME and the front of the second being washed simultaneously at a specific rate of 0.7 gpm/ft^2 and a nozzle pressure of less than 10 psig. Later the valving was changed so that the first ME was washed separate from the second ME. This increased the pressure available for the front wash but not up to the recommended level.

LIMESTONE UTILIZATION

Initial measurements of the limestone utilization in the Duck Creek scrubbers showed that above a reaction tank pH of 5.4, the utilization dropped rapidly. The measurements were made by analyzing a sample of the reaction tank solids for calcium, sulfite, sulfate, and inert material. Utilization is defined as the mole ratio of total sulfur to calcium. It therefore serves as a good indicator of how efficiently the scrubbers are using the limestone fed to the reaction tanks. Including liquid phase calcium and sulfur species results in less than one percent change in the reported utilization. Therefore, all numbers are based on solid phase analysis only.

The low utilizations resulting from operation above pH 5.4 were hurting the FGD systems operation and performance in the following ways:

- reliability -- excess limestone in the slurry contributes to mist eliminator scaling,
- operating costs -- excess limestone could substantially increase annual limestone costs, and
- waste disposal -- presence of the excess limestone in the scrubber blowdown was consuming a significant volume of the disposal pond.

There were two options available for increasing the limestone utilization. First, the reaction tank pH could be controlled at or below 5.4. To do so though, would lower the SO_2 removal efficiency of the scrubbers significantly (see Figure 2) and require higher DBA or magnesium concentrations to make up the difference. The second option would be to modify the ball mill circuit operation to produce a finer and therefore faster dissolving limestone.

During the baseline testing, the product from the grinding circuit was sieved with a 200 mesh and a 325 mesh screen. The mill circuit was originally designed to grind 40 tons per hour of pebble limestone and produce a material of which 90 percent would pass a 200 mesh screen. Sieve results during baseline testing showed that on the average 86 percent passed the 200 mesh screen and 70 percent passed the 325 mesh screen. Review of past sieving data showed that the product varied anywhere from 75 percent passing 200 mesh to 90 percent passing 200 mesh (no data on the percentage passing 325 mesh).

The first and most obvious change examined to improve the grind was to decrease the limestone throughput. Based on calculated SO_2 absorption rates, the mill throughput could be decreased from 40 tons per hour to 20 tons per hour and enough reagent be prepared to keep up with the scrubber demand. After several days of operation at the lower throughput, no measurable change was seen in the product particle size. This suggested that the classifiers were limiting the product fineness.

A series of operational and minor equipment changes were made during June and July with the classifiers and the mill which resulting in a significant improvement in the product particle size distribution. The testing was not designed as a parametric investigation but instead it was conducted as an optimization program. The objective was to produce the finest particle size with only relatively small changes to the existing circuit. The result was a mill circuit that consistently produced a material that greater than 99 percent passed 200 mesh and 90 percent passed 325 mesh.

Although quantification of the effect of each change on the produce particle size is not possible, the following variables were changed in the mill/classifier circuit optimization:

- mill throughput,
- mill slurry solids content,
- level of ball charge in mill,
- top-size of balls in mill charge,
- raw limestone feed size,
- diameter of classifier barrel,
- classifier pressure drop (or throughput),
- solids content in classifier feed, and
- diameter of classifier vortex finder and apex.

The most extensive equipment change made during the program was to replace the 10 inch diameter classifiers with six inch diameter models. Decreasing the diameter of the cyclone barrel theoretically would reduce the size of the largest particle in the product stream by as much as 10 to 30 microns. For comparison, the 200 mesh screen has square openings that are 74 microns in dimension and the 325 mesh screen has openings that are 45 microns square.

The effect of limestone particle size on limestone utilization in the scrubbers was significant. Utilization data taken during the grinding circuit testing was separated into three categories according to limestone particle size. Figure 7 presents the results. The data show that as scrubber pH increases, limestone utilization drops off dramatically. It also shows that at a constant pH, decreasing limestone particle size improved limestone utilization significantly. For example, at pH 5.8, the utilization averaged about 58% with the coarse limestone being fed to the scrubbers, while it

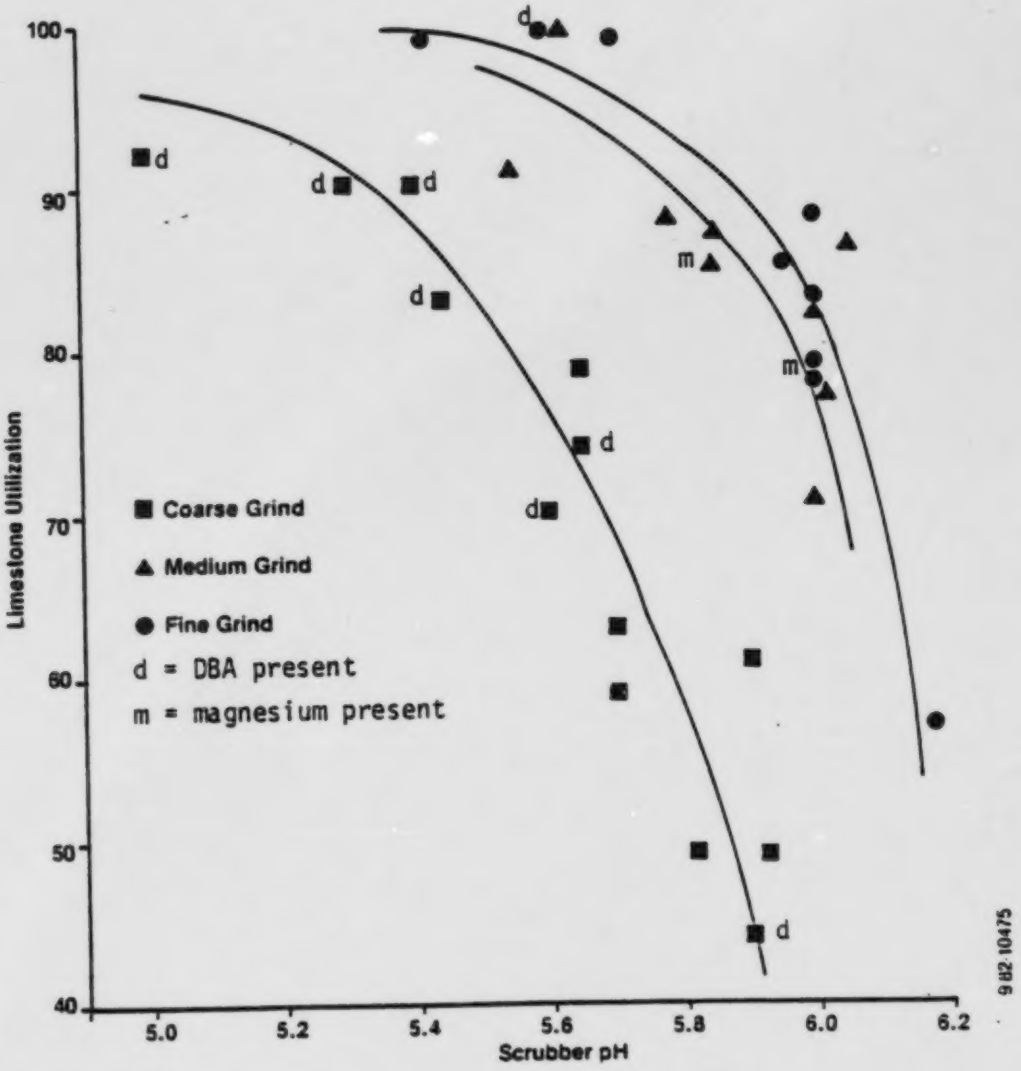

Figure 7. Effect of limestone grind and scrubber pH on limestone utilization

averaged about 92% with the fine limestone. Representative particle size distributions for the three grinds are shown in Table 1. The scatter in the data presented in Figure 7 is primarily due to the variation in the particle size distribution of the limestone within these three categories.

Another important observation made is that the presence of either DBA or magnesium in the reaction tank had no effect on measured limestone utilizations. Samples taken during either DBA or Mg testing are identified in the figure.

TABLE 1. REPRESENTATIVE PARTICLE SIZE DISTRIBUTIONS FOR COARSE, MEDIUM, AND FINE GRINDS

	Particle Size Distribution	
Limestone Size	% Passing 200 Mesh	% Passing 325 Mesh
Coarse	86	73
Medium	94	81
Fine	97	87

As discussed earlier, the improvement in limestone utilization was achieved by relatively minor operational and equipment changes. The primary costs associated with these changes include the capital for the six inch diameter classifiers which totaled less than $10,000 installed, an additional $40,000 per year for electrical power to operate the mill at lower throughputs, and an increase in maintenance costs which can be determined only by time. The benefits by comparison are much more significant. The Duck Creek scrubbers are now being operated at a pH of 5.8. By controlling at this set point the scrubbers not only operate at a higher SO_2 removal efficiency (nearly 82%) compared to previous operation at pH 5.6 but also achieve greater than 90% limestone utilization. In terms of limestone consumption, the Duck Creek scrubbers will use almost 30 million pounds per year less than before. And since this limestone was passing through the scrubbers unused, the waste disposal pond will see an equivalent reduction in material sent to it, thereby extending its useful life. Over a 25-year operating period this would mean a reduction in volume of solid wastes of almost 350,000 cubic yards. This is roughly equivalent to the settled volume of FGD waste produced by the scrubbers in a 5 year period.

COST STUDY

Based on the results of the 1982 test program, a cost study was completed on the economics associated with using DBA and magnesium to improve the Duck Creek scrubbers SO_2 removal efficiency. A total of six options were studied-three each for DBA and magnesium. They investigated the difference between automatic and manual control of additive feed as well as the effect of recycling scrubber blowdown to reduce makeup rates. To simplify the work, a constant reaction tank pH of 5.8 was assumed.

Capital and annual costs were estimated for each case. Based on these costs, a present worth analysis of the various options was performed using the following equation:

$$PW = \frac{S}{(1 + I)^n}$$

where S is the cost for any year n and I equals the discount rate. For this study, CILCo's long term escalation factor of 8 percent was used to escalate both capital and operating expenses. CILCo's long term discount rate of 10.75 percent was chosen for I.

The results of the present worth analysis which is based on a fifteen year period are shown in Table 2. Also shown in the table are the capital investments for each option. The numbers indicate that without the hydroclones to recycle some of the reaction tank blowdown liquor, the magnesium options are not competitive. The reason for this is that much higher liquid phase magnesium concentrations are required than for DBA. Therefore the losses in the waste stream are proportionally higher.

The lowest cost option involves the use of DBA in conjunction with an automatic feed controller and hydroclones for recycle. Both Case 2 and 6 follow next and have about the same present worth cost over the 15 year period at \$2.8 and \$2.4 million each. Case 2 is the DBA option with automatic feed control. The automatic system consists of a microprocessor that receives a signal from the stack gas SO_2 monitor. The microprocessor controls the feed of DBA based on the SO_2 emission level. As the emissions approach the 1.2 lb/10^6 Btu standard the microprocessor would operate a valve to feed DBA to the reaction tanks. As the SO_2 emission level drops below a predetermined value, the microprocessor shuts off the feed. This system would operate the same for the magnesium option. The constant DBA addition option, Case 1, follows at a cost of \$3.0 million.

From an operation and control standpoint, the constant DBA addition option, Case 1, is the easiest to operate. Addition of an automatic controller complicates the system to a small degree but would pay itself off in a short time (approximately two years). Addition of hydroclones would increase the mechanical complexity of the system. Because of the large number of hydroclones which would be required, operating and maintenance related problems would increase as well. Operation of a slaker to prepare the dolomitic lime prior to addition to the reaction tanks makes the magnesium automatic feed/hydroclone recycle option the most mechanically complex system. The operation of a slaker requires close operator supervision. For this reason, purchase of a hydrated lime would be advantageous. With this approach, the material could be fed dry to the reaction tanks or could be mixed with water and pumped as a slurry. Nonetheless, bulk solids handling is inherently more difficult than handling small volumes of an aqueous buffer solution making DBA more attractive from an ease of operation standpoint.

EVALUATION TRIP

A return trip was made to Duck Creek approximately one year following the completion of the 1982 test program. The objective of the trip was to document the performance of the FGD system for comparison to the observations made during the 1982 testing. The results were used to evaluate the ability of the plant personnel to maintain the recommended operating setpoints. They also served as a basis for recommending any changes in controlling the scrubbers to further improve their operation.

TABLE 2. COST ANALYSIS OF DUCK CREEK SO_2 REMOVAL ENHANCEMENT OPTIONS

		Thousands of Dollars (January 1983)	
No.	Case Description	Capital Investment	Cumulative Present Worth[a]
1	Constant DBA Addition	110	3,000
2	DBA Feed Controller	150	2,800
3	DBA Feed Controller Plus Hydroclones	240	1,400
4	Constant Magnesium Addition	140	10,800
5	Magnesium Feed Controller	180	10,000
6	Magnesium Feed Controller Plus Hydroclones	270	2,400

[a] Escalation factor of 8% was used for both fuel and non-fuel related items.
Discount rate assumed was 10.75%.

There were several significant observations:

1) The scrubbers were needing less DBA than anticipated to maintain compliance level SO_2 removal efficiencies.

2) No mist eliminator scaling or pluggage had been seen in the past year.

3) Limestone utilization was not as high at a given pH as had been seen in 1982.

4) FGD system reliability was 96 percent over the period July, 1982, through June, 1983.

5) Limestone product screen analyses consistently showed better than 90 percent passing 325 mesh.

6) A five ton per hour increase in limestone feed to the ball mill had little effect on product size distribution and no noticeable effect on limestone utilization.

The DBA usage and limestone utilization observations were the most interesting. The tradeoffs between low pH/high DBA and high pH/low DBA operation were reviewed. Based on the data collected during the trip, it would be to CILCo's advantage to lower the reaction tank pH setpoints to 5.6 and increase the DBA feedrate. This is because the savings in limestone due to improved utilization at the lower pH more than offsets the increased DBA consumption. Operation below pH 5.6 was not recommended because the pH/utilization relationship begins to flatten below that point (see Figure 7).

The screen analyses showed that the mill circuit was consistently producing a product with over 90 percent passing 325 mesh. Mill throughput was increased from 20 tons per hour to 25 tons per hour to see if this affected the product fineness or limestone utilization in the scrubbers. The amount passing a 325 mesh screen dropped from 92 percent down to 89 percent. The affect on utilization could not be measured if there was any. The resulting 25 percent decrease in mill operating time not only will save on power costs, but should also reduce equipment maintenance.

In general, the control of the FGD systems key setpoints was very good. As mentioned, the limestone mill circuit product fineness could be consistently maintained depending on throughput. Reaction tank pH's were also being controlled within a very small range with good accuracy. Overall, this kept limestone utilization as constant as possible with a single loop (reaction tank) design. Measurements of the solids content in the reaction tanks indicated that the density control system was generally working well. Almost all the measurements fell between 10 weight percent and 13 weight percent with the average being 12 percent solids. Partly as a result of this, the FGD system reliability has averaged slightly above 96 percent since July of 1982 and over 99 percent since January of 1983. At the same time, FGD system operating and maintenance costs have dropped about 30 percent.

TECHNICAL/ECONOMIC FEASIBILITY STUDIES FOR FULL SCALE APPLICATION OF ORGANIC ACID TECHNOLOGY FOR LIMESTONE FGD SYSTEMS

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ABSTRACT

The application of organic-acid buffer enhancement to flue gas desulfurization (FGD) systems is a recent development that has resulted in lowered costs and improved performance for those systems that have adopted its use. A process which uses organic acids as an additive has several advantages over conventional limestone scrubbing systems. These advantages include improved SO_2 removal, decreased limestone consumption, increased system flexibility (e.g., ability to respond to unplanned fluctuations in coal sulfur content), and improved process reliability. This paper summarizes the results of several cost analyses which were performed to evaluate the potential economic benefits of converting operating FGD systems to organic-acid-enhanced limestone scrubbing systems. Also, since the last FGD symposium, two full-scale utility limestone scrubbing systems have converted to organic acid enhanced operations. A summary of the first year of operation for one of those systems—City Utilities Southwest Power Plant (SWPP)—is also included.

BACKGROUND

The ability of a limestone FGD system to remove SO_2 is limited by the absorptive capacity of the slurry liquid that makes contact with the flue gas. Adding an organic acid to the system expands the capacity of the liquid to absorb SO_2 by buffering the pH in the absorber. These are the reactions that take place:

Sulfur dioxide + Water $\rightleftarrows$ Acidity + Sulfites (1)

Organic buffer anion (carboxylate ion) + Acidity $\rightleftarrows$ Organic acid (2)

As the buffer reacts with the acidity formed in the SO_2 absorption reaction, a weak organic acid is formed (Reaction 2) and SO_2 is removed (Reaction 1). The buffer is regenerated in the reaction tank by adding limestone:

$$\text{Organic acid} + \text{Sulfites} + \text{Limestone} \rightarrow \text{Organic buffer anion} + \text{Calcium sulfite solids} + \text{Carbon dioxide} \qquad (3)$$

Adding organic buffers can result in increased SO_2 removal, a decreased risk of chemical scaling, potentially decreased limestone consumption, and potentially reduced mist eliminator fouling because of increased utilization of limestone.

When EPA realized that organic buffers could possibly to used to solve some of the maintenance and cost problems that utilities were experiencing with their lime/limestone FGD systems, the Agency began an extensive research program, designed to provide a thorough evaluation of various organic acid additives and how they can be used to improve lime/limestone scrubbing. It consisted of four phases--laboratory-, pilot-, prototype-, and commercial-scale.

The EPA began its research by sponsoring theoretical and laboratory investigations of a number of different acids. As a result of these studies, adipic acid appeared to be the most promising buffer candidate because of its solubility in water, low volatility, chemical stability, non-toxicity, high availability, and low cost. Also, the region of maximum buffering for adipic acid, between pH 4 and 6, is ideal for limestone FGD systems.

Two full-scale test programs were also sponsored by EPA which demonstrated the technical and economic viability of this approach on operating utility and industrial FGD systems. Results of both of these demonstration programs have been reported in previous FGD symposia.

TECHNICAL AND ECONOMIC ANALYSES OF OPERATING SYSTEMS

Technical and economic analyses were made for four FGD installations to determine the nature and cost of converting each to an organic acid enhanced limestone scrubbing system. The four installations are: City Utilities' Southwest Power Plant, Central Illinois Light Company's Duck Creek Station, San Miguel Electric Cooperative's San Miguel Station, and Big Rivers Electric Corporation's R. D. Green Station. The cost analysis of the R. D. Green Station was not completed in time for this paper. The results will, however, be presented at the FGD Symposium.

The approach used in conducting these process evaluations was basically the same for all systems. First, typical process operating conditions for each system were identified through a series of questionnaires and meetings with plant personnel. Operating parameters such as pH, limestone utilization, liquid to gas (L/G) ratios, and SO_2 removal were obtained from each system to establish baseline process operations and baseline operating costs. Next, technical evaluations were made to identify any process modifications that would be required to convert each system to an organic acid enhanced system, and to establish organic acid makeup requirements. Finally, the capital investment costs for all required process modifications, and the annual operating costs of organic acid enhanced operations were determined. The analysis results for each system are summarized below.

CITY UTILITIES—SOUTHWEST POWER PLANT (1)

City Utilities' (CU) major interest in converting their system to an organic-acid-enhanced limestone scrubbing system was to improve the SO_2 removal performance so that they could comply with the applicable SO_2 removal requirements. Several process alternatives were considered, each of which had the potential for achieving the utility's goal of increased SO_2 removal. The process alternatives considered included: increasing the L/G ratio to the tray tower, adding adipic acid, and adding a mixture of by-product organic dibasic acids (DBA). The DBA used at CU is a by-product of adipic acid production and consists of a mixture of adipic, glutanic, and succinic acids. The higher L/G option represented a capital cost intensive solution; whereas, the adipic acid and DBA options have lower capital costs but higher annual operating costs. A present worth analysis was performed using escalation factors consistent with CU's long range planning activities. The escalation of adipic acid and DBA closely approximated CU's assumed escalation rate of fuel oil, and the escalation of cost of electricity (production costs only) closely approximated CU's assumed coal escalation rate. The discount factor for the analysis was assumed to be equal to the general inflation rate.

The results of this study are shown in Figure 1. Note that the adipic acid option is a lower cost option than increased L/G for about 6 years. The technical uncertainties in making the L/G modification, coupled with the demonstrated adipic acid system flexibility, make the adipic acid option a good choice. However, the DBA option remains the economic choice through more than 15 years. Consequently, City Utilities elected to convert their FGD system to DBA enhanced operations in December 1981. Results of the first year of operation of that system are reported later in this paper.

CENTRAL ILLINOIS LIGHT COMPANY—DUCK CREEK STATION (2)

Central Illinois Light Company's (CILCo's) main interest in considering a conversion to an organic-acid-enhanced limestone scrubbing system was also related to improving the system's performance so that it would achieve compliance with the applicable SO_2 regulations. As with City Utilities, several process alternatives were considered. The options identified for CILCo to achieve compliance included:

- constant DBA addition,
- intermittent DBA feed,
- intermittent DBA plus recycle,
- constant magnesium oxide addition,
- intermittent magnesium oxide feed, and
- intermittent magnesium oxide feed plus recycle.

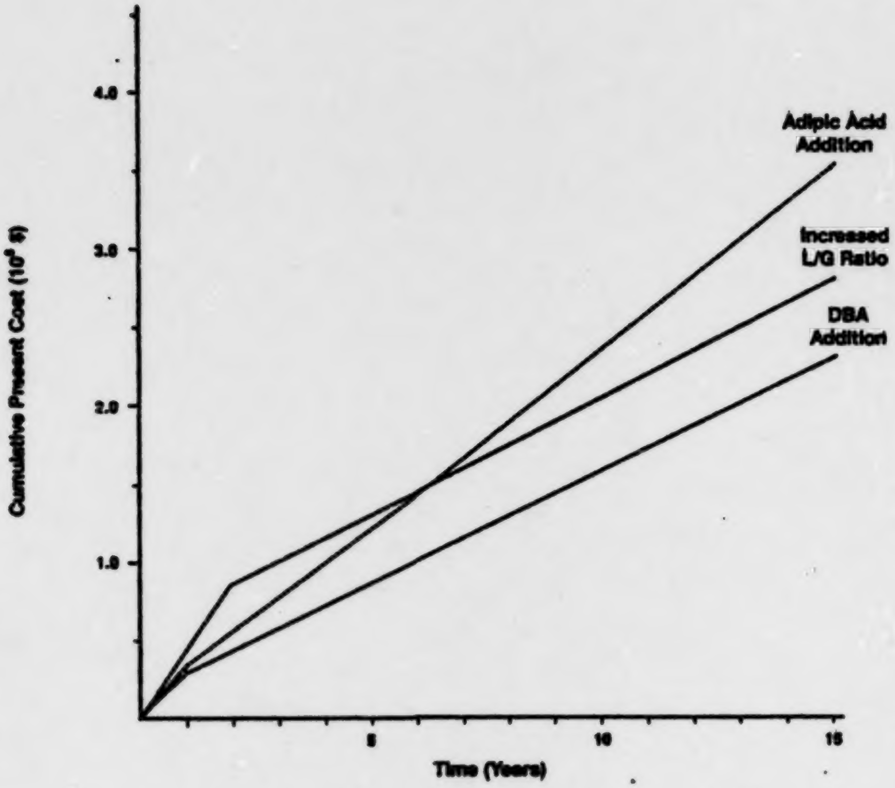

Figure 1. Cumulative Present Cost for Southwest Power Station

The CILCo system is significantly different than the other systems evaluated in that a waste disposal pond is used for solids disposal. In this system, a slurry waste stream of approximately 10 percent solids is sent to the disposal pond, thereby significantly increasing the DBA or magnesium makeup requirements. For this reason, a process alternative, which included the addition of a hydroclone to concentrate the blowdown stream to 40 weight percent solids and thus reduce the additive losses, was also examined.

Results of the present worth analysis over a 15-year period are shown in Figure 2. These results show that the DBA options cost significantly less than the magnesium addition options, except for the case in which a hydroclone is used to reduce the amount of scrubber blowdown sent to the waste disposal pond. Although the process options using the hydroclone showed cost advantages, the use of a hydroclone has not been demonstrated in the field, and thus, technical uncertainties associated with its use exist. For this reason, CILCo has elected to convert their process to a DBA-enhanced limestone scrubbing system with intermittent feed controls so that DBA is added only during high load operations.

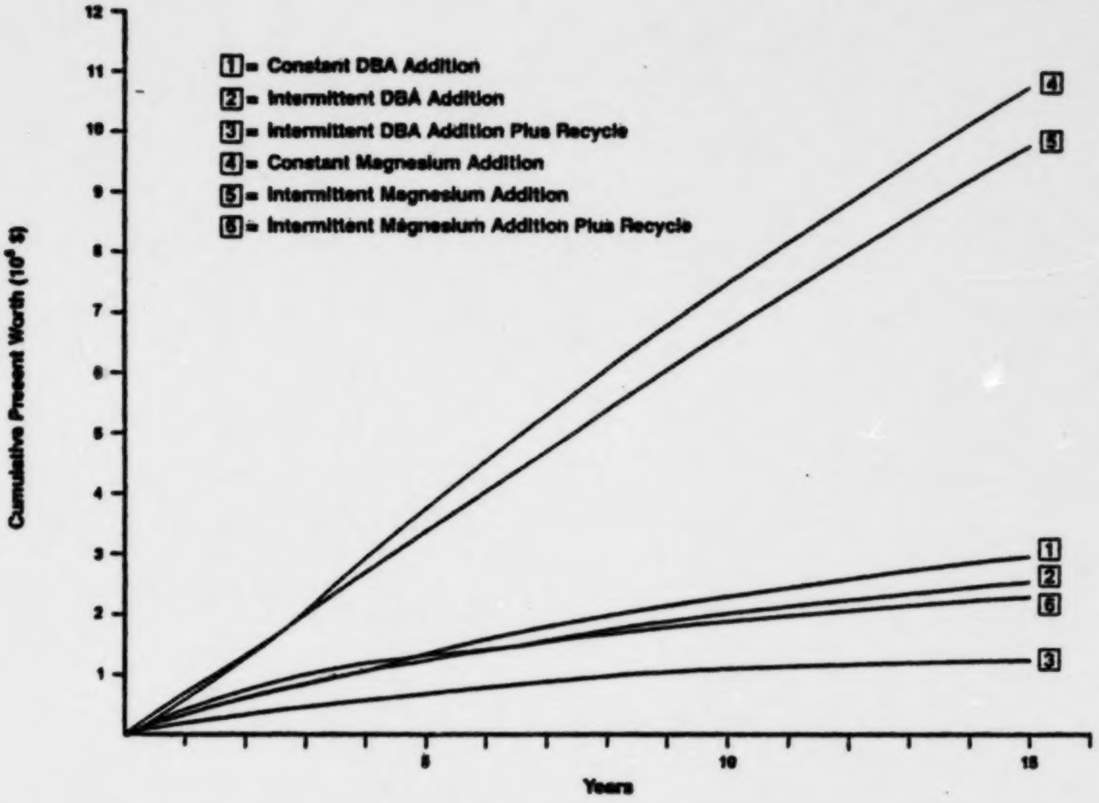

Figure 2. Cumulative Present Worth of Converting CILCo's Duck Creek Station

SAN MIGUEL ELECTRIC COOPERATIVE (3)

San Miguel's interest in converting their system to an organic-acid-enhanced system stemmed primarily from a desire to reduce their FGD operating costs rather than from regulatory pressures. Normal operating conditions at San Miguel required the use of fairly high limestone reagent ratios in order to meet compliance. Reagent ratios on the order to 1.2 to 1.3 are routine, and as high as 1.4 have been measured. Besides the obvious costs associated with excess limestone use, San Miguel also experienced chemical scaling in their absorber tower which was probably related, at least in part, to the excess limestone present in the system.

This paper presents the results of a cost analysis to determine the potential cost savings associated with adding DBA to achieve an SO_2 removal level of 85 percent. Due to the uncertainties in the DBA feed requirement and the current limestone reagent ratio at San Miguel, a range of annual operating and maintenance (O&M) expenses was used in the cost analysis. These expenses reflect the incremental change in O&M expenses from the current FGD operation. Maximum O&M expenses correspond to the maximum estimated DBA requirement (400 ppm) and the minimum limestone savings (which result if San Miguel's current limestone reagent ratio is 1.2). Minimum O&M expenses correspond to the minimum estimated DBA requirement (300 ppm) and the maximum limestone savings (which occur if San Miguel's current limestone reagent ratio is 1.3). Average O&M expenses assume a DBA feed requirement of 350 ppm and a current limestone reagent ratio of 1.25.

A present worth analysis was performed based upon San Miguel's long term escalation and market interest rate assumptions and is presented in Figure 3. Taking into account the initial capital investment for the DBA system, the cost savings for the minimum and average O&M expenses were estimated over a 15-year period at $1.5 and $0.4 million in terms of present worth. Payback periods were calculated to be 1.2 and 2.5 years from the the beginning of 1983, respectively. Similarly, if the maximum operating expenses are realized, the DBA system will cost San Miguel $657,000 over the 15-year present worth period.

Note that no credit for increased reliability was assumed in any of the cost calculations. For San Miguel, which has experienced chemical scaling problems, the savings could be very significant. Worst case assumptions predict a cumulative cost for a DBA system of $657,000 over a 15-year period. If DBA allows San Miguel to remain on line for only 2 extra days during the 15 years, it will have saved the total present worth cost of the DBA system for the entire 15-year period.

Because of the potential cost savings and system performance improvements associated with the use of a DBA system, San Miguel has decided to conduct a test program to verify the results of this engineering evaluation. Field testing of DBA-enhanced limestone scrubbing operations is to be conducted in late 1983.

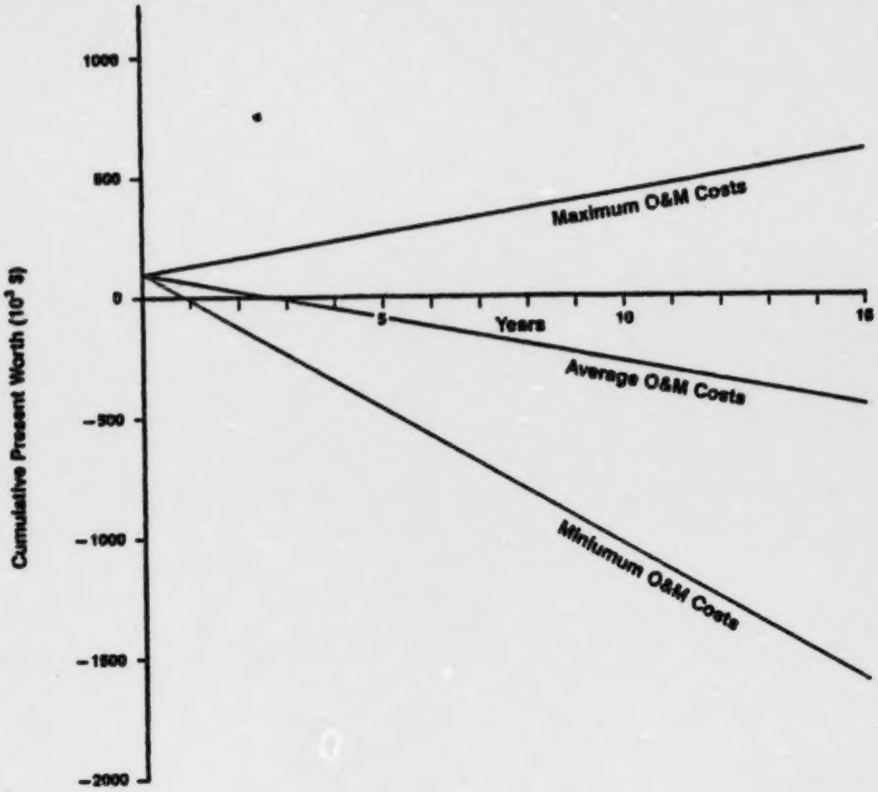

Figure 3. Cumulative Present Worth of Converting San Miguel's FGD System to DBA Assisted Operations

RESULTS OF THE FIRST YEAR OF COMMERCIAL OPERATION OF AN ORGANIC-ACID-ENHANCED FGD SYSTEM

In 1980, Radian Corporation and City Utilities entered into an agreement to participate in a demonstration project which was sponsored by U.S. EPA's Industrial Environmental Research Laboratory at Research Triangle Park, NC. Adipic acid and dibasic acid (DBA) were tested as buffering agents in the Southwest Unit 1 limestone scrubber to determine their effectiveness in increasing utilization and SO_2 removal. This testing was conducted during 1981 and verified the effectiveness of organic acid buffers in a commercial limestone scrubber.

It was quite important to CU officials that improved scrubber operating efficiencies be obtained immediately. In late 1980, Region 7 of the U.S. EPA cited the operation of Southwest Power Station Unit 1 for failure to comply with flue gas emission regulations. By October 1981, upon completion of Radian's testing, negotiations were ongoing with EPA in an attempt to resolve the dispute. In efforts to reach an agreement on a "Consent Decree" City Utilities agreed to use a material, such as adipic or dibasic acid, as an additive to enhance the reactivity of the $CaCO_3$ and thus increase the SO_2 removal from the flue gas. City Utilities embarked immediately on a plan to implement the use of DBA (4). This section of the paper describes the results of the first year of DBA operations.

DBA FEED SYSTEM

A temporary DBA feed system was installed in late December 1981 to add the DBA solution on a continuous basis. This temporary system used a 26,000 ℓ (6,000 gal) stainless steel tanker trailer equipped with thermal jacketing as a storage vessel. Two residential-sized electrical water heaters provided a source of hot water at ∿ 60°C (140°F) for circulation through the thermal jacket of the truck. The heater return lines were run in parallel with the DBA feed line through a 10 cm (4-in.) PVC conduit in order to prevent feed line cooling. Maintaining high DBA temperatures was a major concern because DBA tends to easily precipitate out of aqueous solution. As an example, a 20 percent solution of DBA will begin to show crystal formation at 22°C (72°F); higher concentrations require even higher temperatures to maintain solubility. The entire temporary system was assembled on-site and tied into the scrubber system in a matter of only a few days.

Initially, DBA was pumped from the tank truck to the ball mill sump. The feed rate was set manually. After the DBA was mixed with the freshly ground limestone slurry from the ball mills, the mixture was pumped to the limestone storage tank. Since the limestone handling equipment was common to both scrubber modules, both scrubbers operated with the same DBA concentration. Late in September, the DBA feed system was modified to allow DBA to be fed separately into the reaction tank of either scrubber. (5)

SO_2 REMOVAL

Results of a year's worth of testing using DBA as an FGD additive were consistent with those anticipated based on the demonstration project conducted in 1981. On the whole, FGD performance was enhanced to such an extent that Southwest Power Station consistently operated in compliance with the 1.2 lb $SO_2/10^6$ Btu (516 ng/J) federal emission limitation on a day-to-day basis. The few emission exceedances experienced were generally associated with start-ups, shutdowns, and other periods of allowable operational curtailment.

Figure 4 displays a monthly history of FGD performance, expressed as overall SO_2 removal efficiency, for the 1982 calendar year. It can be seen that efficiencies were consistently greater than the approximately 82 percent removal generally required at the Southwest Station to maintain compliance. For comparative purposes, Figure 4 also displays the performance history for calendar year 1980, wherein removal efficiencies were erratic and averaged only 26.0 percent. Figure 5 shows these data in terms of average monthly emission rates. This figure illustrates the significant improvement in the ability of the system to comply with the regulatory emission standard after the conversion to a DBA enhanced system. (6)

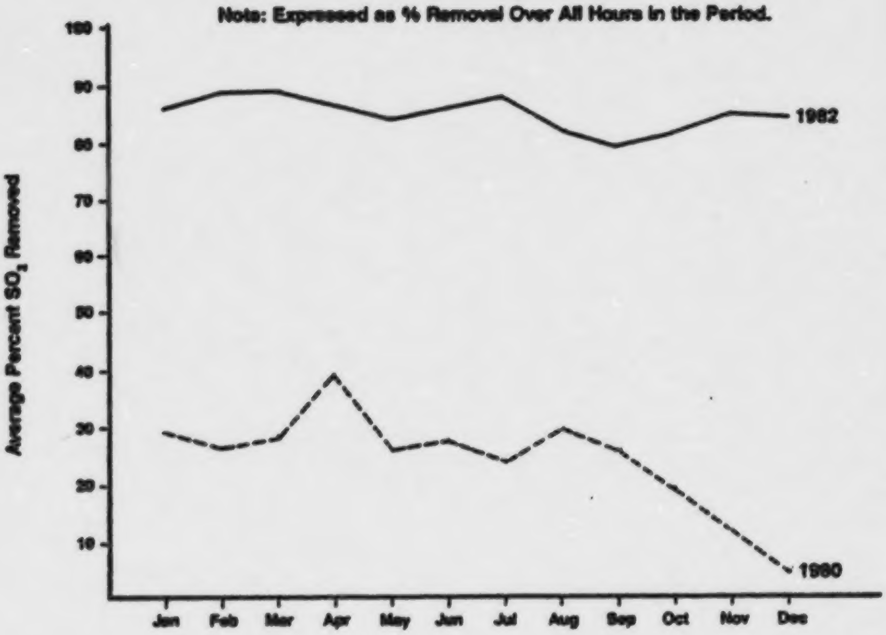
Note: Expressed as % Removal Over All Hours in the Period.

Figure 4. Comparison of Southwest Power Station SO_2 Removal Efficiencies, 1980 vs. 1982

The addition of DBA also improved the system's ability to respond to various operational changes. During a 2-week period in March, the coal supplied to the system contained 4.5 percent sulfur instead of the 3.5 percent sulfur level for which the system was designed. During this period, the unit was able to maintain an in-compliance status by increasing the amount of DBA feed to the system. (7)

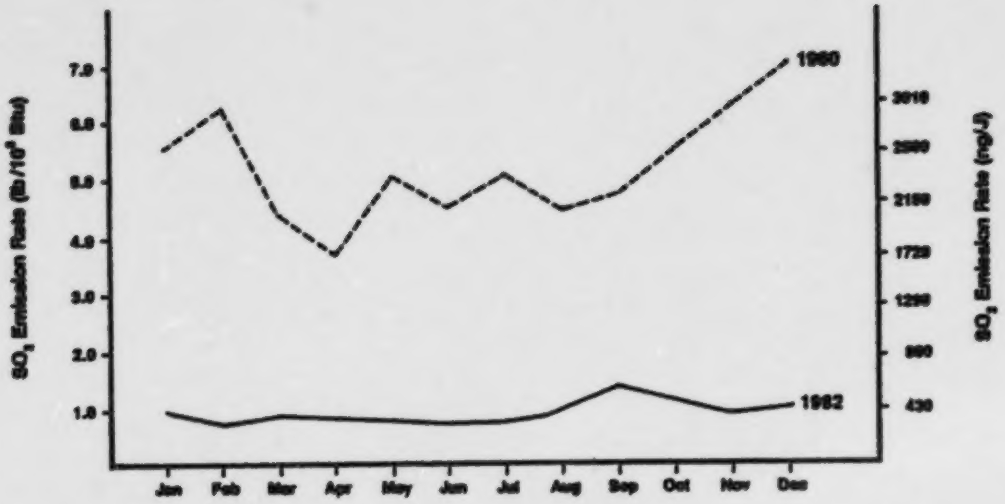

Figure 5. Comparison of Southwest Power Station Monthly Average SO_2 Emission Rates, 1980 vs. 1982

RELIABILITY

The overall reliability of the system was also increased significantly in 1982 in comparison with other years. Figure 6 shows the monthly average FGD reliability for 1982 compared with the average reliability for 1979 and 1980. (8) This reliability increase may have resulted in part from DBA addition, but was also related to a number of process improvements made during 1981 and 1982, such as the conversion from turbulent contact absorption towers to tray towers.

CONTINUOUS MONITORS

The performance of the continuous emission monitoring system (CEMS) used by CU to determine their SO_2 emissions was also evaluated to determine the ability of the CEMS to gather the type of data required by the 1979 revision to the utility New Source Performance Standard. To meet EPA requirements for data capture, at least two data points must be taken each hour for a minimum of 18 hours each day for 22 days per month. The goal of the CEMS at SWPP was to collect at least four data points per hour for 24 hours.

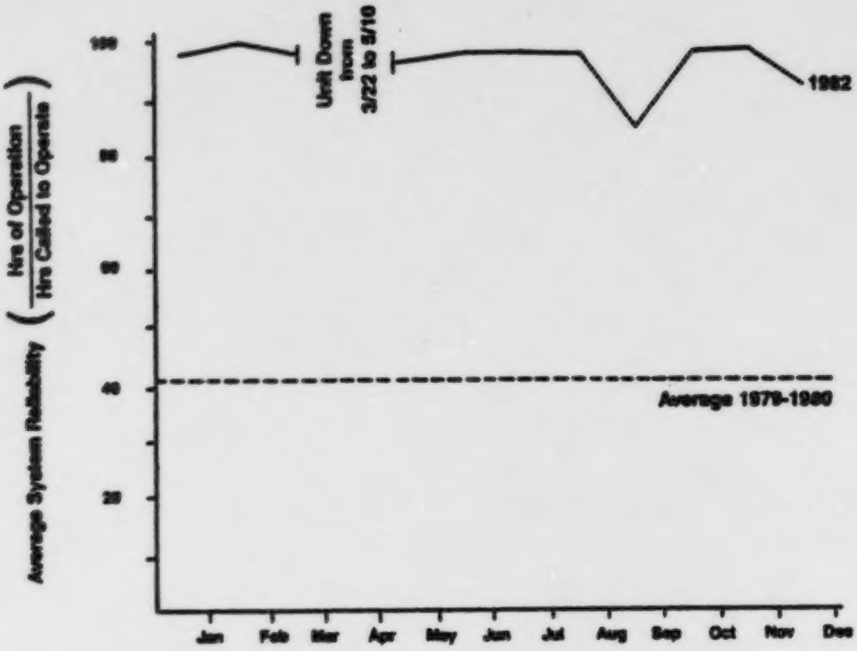

Figure 6. Comparison of Southwest Power Station Process Reliability, 1982 vs. Average Reliability for 1979-1980

Monitor reliability data were collected during the last 6 months of 1982. During this time, the CEMS proved to be very reliable, capturing in excess of 97 percent of the available data for each scrubber module. Table 1 presents the data capture percentage for each module by month and for the program as a whole. The lowest data capture efficiency was 94 percent during July on the S-1 Module. The majority of the downtime encountered by the CEMS was caused by power interruptions. (9)

TABLE 1. SUMMARY OF CONTINUOUS MONITOR RELIABILITY

	Module S-1	Module S-2
June	98.9	100.0
July	94.0	96.8
August	95.9	96.1
September	99.0	98.4
October	100.0	99.9
November	97.4	97.6
December	96.2	100.0
Average	97.3	98.4

DBA CONSUMPTION

The DBA consumption rate increased significantly during the last half of 1982 to where it was nearly twice that observed during the 1981 demonstration tests. The increased DBA consumption was due to two main factors: changes in the recycle slurry pH and deterioration of the slurry recycle pumps. Slurry pH was decreased in 1982 from 5.4 to 5.1 to improve limestone utilization in an attempt to improve mist eliminator reliability. (Mist eliminator performance has been greatly improved by this technique and by revising the system water balance to wash with greater amounts of fresh water.) While the lower pH does improve limestone utilization, a higher concentration of DBA is required to achieve the desired SO_2 removal. (10)

The deterioration in the recycle slurry pumps over time resulted in an actual L/G ratio less than indicated. In order to maintain the same SO_2 removal at a decreased L/G, the concentration of DBA was increased. The pumps underwent a complete overhaul at the plant's scheduled outage in early 1983 to correct this problem.

FUTURE ACTIVITIES

Several operational improvements were implemented during the 1983 extensive maintenance outage. An example is a limestone classification system and a permanent DBA addition system. Based on the experiences of the Duck Creek Station of the Central Illinois Light Company in their DBA program, it was found that by improving the efficiency of the limestone classification system they could reduce DBA consumption on the order of magnitude of 10 percent. During a short period of classifier testing during November 1982, similar, yet inclusive, results on DBA usage were obtained by CU.

With the improved operator control over the DBA feed system and the direct feed to each hold tank, a much quicker chemical response will be obtained which should allow consistent SO_2 emissions rates at or slightly below the required 1.2 lb/10^6 Btu (516 ng/J) level. An increased reduction of DBA usage is projected as the operator exercises more precise system control. When considering the beneficial effects of the classifier system as well, overall DBA consumption should be reduced significantly as compared to 1982 values.

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6. Reference 4.

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9. Reference 5.

10. Reference 5.

SESSION 9: WASTE DISPOSAL/UTILIZATION

Chairman: James D. Kilgroe
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Full-Scale Field Evaluation of Waste Disposal from Coal-Fired Electric Generating Plants

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To be presented at the
Eighth Symposium on Flue Gas Desulfurization
New Orleans, Louisiana
November 1-4, 1983

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ABSTRACT

This paper summarizes results of a 3-year study of current coal ash and flue gas desulfurization (FGD) waste disposal practices at coal-fired electric generating plants. The study was conducted by Arthur D. Little, Inc., under EPA contract 68-02-3167, and involved characterization of wastes, environmental data gathering, evaluation of environmental effects, and engineering/cost evaluations of disposal practices at six selected sites in various locations around the country. Results of the study are expected to provide technical background data and information to EPA, state and local permitting officials, and the utility industry for implementing environmentally sound disposal practices.

Data from the study suggest that no major environmental effects have occurred at any of the six sites; i.e., data from wells downgradient of the disposal sites indicate that waste leachate has resulted in concentrations of chemicals less than the EPA primary drinking water standards. A generic environmental evaluation based on a matrix of four waste types, three disposal methods, and five environmental settings (based on climate and hydrogeology) shows that, on balance, technology exists for environmentally sound disposal of coal ash and FGD wastes for ponding, interim ponding/landfilling, and landfilling. For some combinations of waste types, disposal methods, and environmental settings, mitigation measures must be taken to avoid adverse environmental effects. However, site specific application of good engineering design and practices can mitigate most potentially adverse effects of coal ash and FGD waste disposal. Costs of waste disposal operations are highly system and site specific.

Introduction

This study--of current coal ash and flue gas desulfurization (FGD) waste disposal practices at coal-fired power plants--was conducted for the U.S. Environmental Protection Agency (EPA) by Arthur D. Little, Inc. (ADL) under EPA contract 68-02-3167. The study involved characterization of wastes, environmental data gathering, evaluation of environmental effects, and engineering/cost evaluations of disposal practices at six selected sites at various locations around the country. Results of the study are expected to provide the technical background data and information needed to help EPA determine the degree to which disposal of these wastes needs to be managed in order to protect human health and the environment. The study results will also assist EPA in preparing a report to Congress required under the 1980 Amendments to the Resource Conservation and Recovery Act (RCRA), and should provide useful technical information to federal, state, and local permitting officials and utility planners on methods for environmentally sound disposal of coal ash and FGD wastes. Results of this 3-year effort are summarized in this paper.

Background on Waste Generation/Disposal Methods

Coal-fired power plants using conventional combustion technology generate two major categories of waste materials. Coal ash (fly ash, bottom ash, or boiler slag) and FGD wastes are generated in large amounts relative to other wastes generated at these plants and, therefore, are usually referred to as "high volume wastes." Numerous other wastes, generated in smaller quantities, are associated with other processes or maintenance operations in a power plant such as coal pile runoff, boiler blowdown, cooling tower blowdown, water treatment wastes, maintenance cleaning wastes, general power plant trash, and plant sanitary wastes. This project primarily focused on the high volume wastes.

Fly ash from coal-fired utility boilers is collected by mechanical collectors and/or electrostatic precipitators, fabric filters, or wet scrubbers. By late 1982, approximately 103,000 MW of coal-fired generating capacity--operational units, units under construction, and units at various stages of planning--had been committed to FGD systems. Flue gas desulfurization can be accomplished by nonregenerable throwaway systems, which result in FGD wastes, or by regenerable systems, which produce a saleable product (sulfur or sulfuric acid). Operational nonregenerable FGD systems are currently predominated by wet scrubbing technology; however, some dry FGD scrubbing systems were becoming operational in 1982-1983. The principal types of systems used in utility power plants are those based on direct limestone, direct lime, alkaline fly ash, dual alkali, and lime- or sodium-based dry FGD systems.(1)

Some projections on the generation of coal ash and FGD wastes [together, these are designated as flue gas cleaning (FGC) wastes] in the U.S. are presented in Table 1. Most of the coal ash and all of the FGD wastes generated are sent to disposal.(2,3) Considering the expected increase in coal consumption in the U.S., this is likely to be the case for many years. Utilization of FGC waste is expected to grow but at a slower rate than FGC waste generation. A significant fraction of the total coal ash generated is used for such purposes as soil stabilization, ice control, and as ingredients in cement, concrete, and blasting compounds; however, there is currently no utilization of FGD wastes in the U.S. On balance, disposal will continue to be the major option for FGC waste management in the U.S. for the foreseeable future.

Currently, all FGC waste disposal is on land. At-sea disposal may be a future alternative if it can be practiced under environmentally and economically acceptable conditions. The principal methods of disposal on land are: ponding, landfilling (including disposal in surface mines), and interim ponding followed by landfilling. Table 2 presents data on current practices based on data obtained on 176 plants.

Ponding of FGC waste is more widely practiced than any other disposal method. Ponding can be employed for a wide variety of coal ash and FGD wastes including chemically treated FGD wastes. Ponds can be designed based on diking or incision, but the construction of dams or dikes for ponds is usually expensive. In the future, particularly if chemical treatment of FGD wastes is widely practiced, ponding will probably be limited to those sites that would involve minimal construction of dams or dikes. One exception could be a special case of wet ponding--FGD gypsum "stacking." In this case, gypsum slurry from a forced oxidation system would be piped to a pond and allowed to settle and the supernate recycled. Periodically the gypsum would be dredged and stacked around the perimeter of the pond, thus building up the embankments.

Landfilling of FGC waste is also widely practiced, and can involve one or more of a variety of handling operations prior to the disposal operation. For example, bottom ash is almost always sluiced from the plant, so it must be dewatered (e.g., via hydrobins) before it is transported. Dewatering must also be applied to fly ash that is sluiced from the plant or is wet-scrubbed from the flue gas--with or without significant quantities of SO_2. Wet FGD waste must also be dewatered via thickening, vacuum filtration, and, if necessary, blending with dry fly ash for stabilization or other chemical treatment ("fixation") additives such as lime. On the other hand, fly ash slated for landfill is typically transported directly from the plant in a dry state, with only enough moisture added as required for dust control and compaction in the landfill. Wastes from a spray dryer FGD system can also be transported directly; during this project, commercial operation of these systems on utility boilers was just beginning.

TABLE 1. PROJECTIONS OF FGC WASTE GENERATION BY UTILITY PLANTS IN THE UNITED STATES (1980-1995)

	Waste Generation (10^6 Metric tons/yr)		
Waste Type	1980	1985	1995
Coal Ash[a]	62.4	83.2	110.0
FGD Wastes[b]	8.6	26.9	48.6
TOTAL	71.0 (78.3)[c]	110.1 (121.4)[c]	158.6 (174.8)[c]

[a] Coal ash quantities are shown on a dry basis.

[b] FGD waste quantities are shown on a wet basis (50% solids).

[c] 10^6 tons/year.

Source: Reference 2

TABLE 2. CURRENT FGC WASTE DISPOSAL METHODS UTILIZED AT UTILITY COAL-FIRED POWER PLANTS IN THE U.S. (Data Base: 176 Plants ≥ 200 MW)[a]

	Number of Plants		
Type of Waste	Pond[b]	Landfill[c]	Interim Pond/Landfill[c]
Fly ash only	18	46	6
Bottom ash only	29	13	29
Combined fly and bottom ash	69	9	16
FGD waste only	5	-	-
Mixed fly ash and FGD waste	7	7	-
Mixed bottom ash and FGD waste	1	-	1
Mixed fly ash and FGD waste (stabilized)	2	6	-
Mixed fly ash, bottom ash, and FGD waste	2	1	1

[a] Coal-fired plants on which data were available (≥80% of their power generated from coal in 1977) which have generating capacities ≥200 MW with the exception of four plants employing FGD systems. Figures represent the number of plants at which each waste type/disposal method is practiced. (Note: Many plants utilize more than one method.)

[b] Includes direct ponding and interim/final ponding methods.

[c] Includes managed and unmanaged fills and mine disposal.

Source: Reference 4

In a landfill disposal site, the wastes are spread on the ground in 0.3 to 1 m (1 to 3 ft) lifts and compacted by wide track dozers, heavy rollers, or other equipment. Layering proceeds in 0.3 to 1 m lifts in segments of the site. The ultimate height of a disposal fill is site specific, but can range from 10 m (30 ft) to as high as 76 m (250ft). A properly designed and operating dry impoundment system can potentially enhance the value of the disposal site after termination or at least permit post-operational use.

Mine disposal is a variation of landfilling that is receiving increased attention. Surface coal mines, particularly those serving "mine-mouth" plants, offer the greatest capacity and economic attractiveness for disposal of wastes from power plants.(5,6) Since the quantity (volume) of FGC wastes produced is considerably less than the amount of coal burned, many mines would have the capacity for disposal throughout the life of the power plant. Several plants, particularly in the Plain States (e.g., North Dakota), have practiced this disposal method in recent years.

Interim pond/landfill has been an important waste disposal method in the past, but is likely to decline in importance in the future, particularly since dry ash handling and disposal is being more widely practiced.

SITE SELECTION AND TEST PLAN PREPARATION

Candidate Site Selection Process

The overall objective of the candidate site selection process was to evaluate available data on coal-fired power plants and recommend a number of candidate and backup sites. The selection process consisted of two steps:

First, the contiguous 48 states were divided into 14 physiographic regions, and the plants in each region were screened to develop a list of plants that would be suitable for consideration as candidate and backup sites. A total target of 25 to 30 sites, including 18 candidate and 7 to 12 backup sites, was desired. Based on an assessment of present and future FGC waste disposal practices, a preliminary distribution of the targeted number of candidate sites in each region was agreed upon. In screening selections, the investigators remained cognizant of the targeted number in each region, but were not absolutely limited by that number. The attempt was to choose desirable plants in as many regions as possible. The list of plants in all the regions that came through this filtration process amounted to 26.

Second, these 26 were then ranked in iterative group discussions leading to the nomination of 18 as candidate sites and the remainder as backup sites.

Final Selection Process

The candidate and backup sites were then subjected to a more detailed evaluation. These evaluations included one or more detailed site visits by engineering, environmental, and hydrogeologic specialists assigned to the project. Their findings, together with mid-course evaluations that were continuously taking place, supported an iterative process that resulted in the selection of the final six sites. Table 3 provides overall information on the final six sites that were selected for evaluation under this project; Figure 1 indicates the site locations.

Test Plan Preparation

Detailed test plans providing background information on each of the sites, together with a description of the proposed program of site development, physical and chemical sampling, and analysis and engineering/cost assessments were developed. The test plans were reviewed by EPA and the utility involved, and their comments were incorporated. The finalized test plans guided the work at each site.

SITE DEVELOPMENT AND PHYSICAL TESTING

After approvals from the utility and, in some cases, from state regulatory agencies, site development was begun. Site development and physical testing were governed by procedure manuals(4) developed for this project. The activities involved in site development included the drilling of borings; excavating test pits; collecting waste, soil, and water samples; conducting field permeability tests; installing ground water monitoring wells and piezometers; and documenting each activity. These activities took place at each of the six sites in time periods of 2 to 4 weeks. Table 4 indicates the timing under which the six sites were developed and the extent of the activities at each site. The table also gives the number of physical tests performed; i.e., laboratory soil classification and permeability tests on waste samples from the sites. Preliminary water balances were also developed for each site.

CHEMICAL SAMPLING AND ANALYSIS

At each site, a program of chemical sampling and analysis was undertaken. This program included characterization of waste, water, and soil samples obtained from site development, and ground water well and (in some cases) surface water samples subsequently obtained from a series of visits scheduled to correspond to relatively wet, relatively dry, and intermediate periods for each site. Table 5 shows a summary of the sampling and analysis program.

TABLE 3. WASTE DISPOSAL SITES SELECTED FOR EVALUATION

Plant (Utility)	Location State (County)	Nameplate Generating Capacity (MW)	Startup Date (mo/yr) Plant (FGD)	Waste Site Under Study Waste Type	Waste Site Under Study Disposal Method	High Priority Issues Under Study Ground Water Quality	High Priority Issues Under Study Surface Water Quality	High Priority Issues Under Study Employment of a Potentially Mitigative Practice
Allen (Duke Power)	N. Carolina (Gaston)	1155	-/57	Combined fly and bottom ash	Pond (UL)	X	X	X
Elrama [Duquesne Light (waste disposal by Conversion Systems, Inc.)]	Pennsylvania (Washington)	510	6/52 (10/75)	Stabilized FGD waste	Landfill (UL; offsite)	X	X	X
				Combined fly and bottom ash	Landfill (UL; offsite)			
Dave Johnston (Pacific P&L)	Wyoming (Converse)	750	-/59	Fly ash	Landfill (UL)	X	-	X
Sherburne County (Northern States Power)	Minnesota (Sherburne)	1458	5/76 (5/76)	Fly ash/FGD	Pond (CL)	X	-	X
Powerton (Commonwealth Edison)	Illinois (Tazewell)	1786	-/72	Combined fly and bottom ash	Landfill (AL)	X	X	X
Smith (Gulf Power)	Florida (Bay)	340	6/65	Combined fly and bottom ash	Pond (UL)	X	X	X

UL = Unlined.
CL = Clay-Lined.
AL - Artificially Lined (Poz-O-Tec).

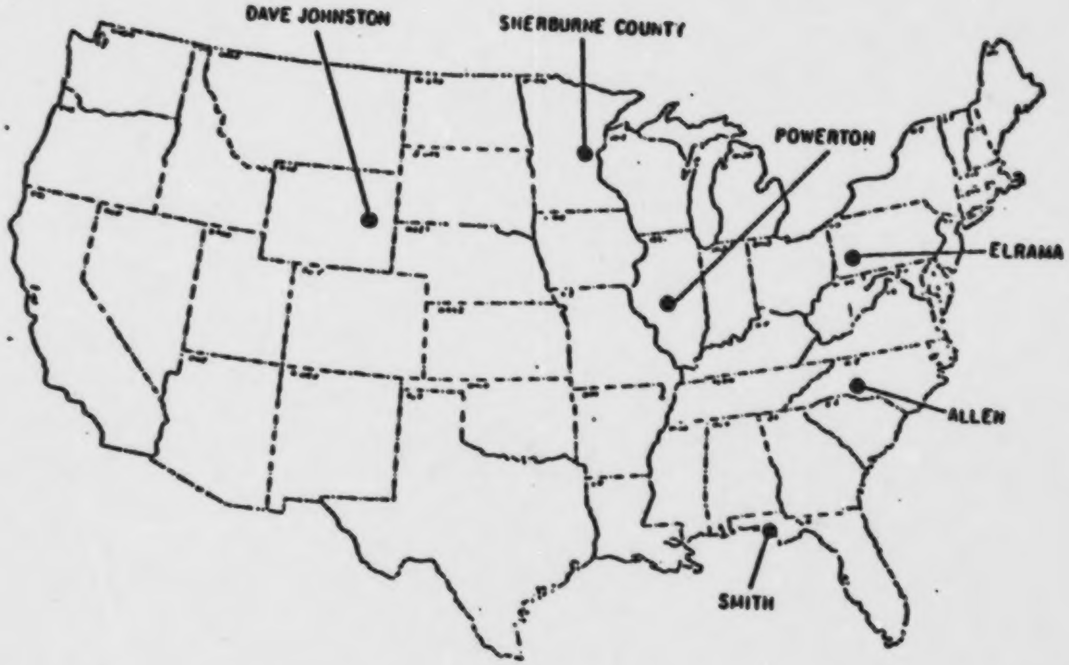

FIGURE 1. LOCATION OF WASTE DISPOSAL SITES SELECTED FOR EVALUATION

TABLE 4. SUMMARY OF SITE DEVELOPMENT/PHYSICAL TESTING

Plant	Date Development Completed (mo/yr)	Number of				Number of Laboratory Physical Tests	
		Borings	Wells	Test Pits	Soil Samples	Unified Soil Classification Series (USGS)	Permeability
Allen	01/81	20	20	2	152	18	4
Elrama	03/81	20	16	4	199	17	13
Johnston	05/81	14	12	10	154	12	7
Sherco	08/81	13	11	-	178	20	6
Powerton	11/81	11	9	1	112	30	8
Smith	12/81	25	24	-	146	15	8

TABLE 5. SUMMARY OF CHEMICAL SAMPLING AND ANALYSIS PROGRAM

Site	Samples[a]		Analyses[b]				
			Trip 1	Trips 2, 3, and 4			
	Trip 1	Trips 2, 3, and 4	ICAP	IC	As/Se	Field Data	Other
Allen[c]	wells	wells and surface waters	X	X	X	X	
	ash solids		X		X		
	interstitial liquors		X	X	X		
	soils		X		X		
Elrama	wells	wells, lysimeters, surface waters	X	X	X	X	
	waste solids		X		X		X[d]
	soil		X		X		
	waste extracts			X			
Sherco	wells	wells and surface liquors	X	X	X	X	
	waste interstitial liquors		X	X	X		
	waste solids		X				
	liner solids		X				
	liner liquor		X	X			
	soil solids		X				
	soil extracts		X				
Smith	liquids	wells and surface waters	X	X	X	X	
	waste solids		X				
	interstitial waste liquors		X	X			
	soil		X				
	soil liquors		X	X			
Powerton	wells	wells and surface waters	X	X	X	X	
	waste solids		X				
Dave Johnston	wells	wells and surface waters	X	X	X	X	
	waste solids		X		X		
	waste extracts			X			
	soils		X				

[a] Samples obtained during site development and subsequent sampling and analysis trips.

[b] Analyses performed are abbreviated as follows:

ICAP - Ag, Al, B, Ba, Be, Ca, Cd, Co, Cr, Cu, Fe, K, Mg, Mn, Mo, Na, Ni, P, Pb, Si, Sr, Th, Ti, V, Zn, Zr. (Does not include B, Ba, and Si for solids.)
IC - F^-, Cl^-, NO_3^-, $SO_4^=$, Br^-, PO_4^{3-}.
As/Se - either or both on selected samples.
Field Data - ground water level, pH, dissolved oxygen, conductivity, temperature.

[c] Other samples were obtained (boiler cleaning wastes). Analysis was limited to ICAP, pH, and bromate.

[d] Includes solids characterization for SO_4^{2-}, total oxidizable sulfur, slurry pH, acid insolubles.

Chemical samples are subjected to several types of analyses: ion chromatography (IC) for six anions, inductively coupled argon plasma emissions spectroscopy (ICAP) for 26 metals; and atomic absorption spectroscopy (AA) for selected metals. As shown in Table 5, these types of analyses were performed on a mix of solid and liquid samples for each site. In addition, a limited number of experiments were performed to assess the attenuative capacity of various soils obtained at the sites. Furthermore, during the initial phase of this project, 23 grab samples of wastes from 18 plants were obtained and analyzed using the EPA Extraction Procedure (EP); a summary of results from these tests is given in Table 6. Further details on these tests, as well as results of radioactivity measurements, are included in Reference 4.

SITE-SPECIFIC ENVIRONMENTAL EVALUATIONS

The data and information from site development and sampling/analysis were subjected to environmental effects evaluation throughout the project. The individual site evaluations were developed by a series of five sequential steps.

First, a review and evaluation was made of available background information on the disposal operation and its environmental setting. Second, present disposal-related water quality effects were identified and described based on evaluation and measured information developed in this project. Third, apparent cause/effect relationships were hypothesized to explain the findings at the sites. Fourth, potential future ranges of water quality effects were considered to the extent that suitable data were available. Finally, industry-wide implications of the findings at the individual sites were considered in the generic assessment, discussed later in this paper.

Environmental evaluation of all six sites has generated a significant amount of data and information. The following general items can be reported:

1. Data suggest that no major adverse environmental effects have occurred at any of the sites. For example, data from wells downgradient of the disposal sites suggest that the contribution of waste leachate to the ground water has resulted in concentrations of chemicals less than the primary drinking water standards established by EPA.
2. The results from the sites are internally consistent. In other words, the analyses of samples taken on different dates at the same locations are very similar.
3. The total integrated evaluation of data from site development, site water balances, physical testing of wastes samples, and chemical sampling and analysis is providing a large significant data base to explain many of the environmental effects that can result from coal ash and FGD waste disposal.

A brief account of the results at each of the sites is presented below.

TABLE 6. SUMMARY OF RESULTS OF EXTRACTION PROCEDURE (EP) TESTS OF 20 FLY ASH AND 3 FGD WASTE GRAB SAMPLES

Metal	Overall Range Observed, μ g/l		Interim Primary Drinking Water Standards[a], μ g/l	Ratio of Range Observed to Standards	
	Fly Ash	FGD Waste		Fly Ash	FGD Waste
Arsenic	<2 - 410	<2 - 65	50	<0.04 - 8.2	<0.04 - 1.30
Barium	<100 - 700	<150 - 230	1000	<0.1 - 0.7	<0.15 - 0.23
Cadmium	<2 - 193	<2 - 20	10	<0.2 - 19.3	<0.2 - 2
Chromium	<8 - 930[b]	<11 - 26[b]	50[b]	<0.16 - 18.6	<0.22 - 0.52
Lead	<3 - <36	<5	50	<0.06 - 0.72	0.1
Mercury	<2	<2	<1	<1	<1
Selenium	<2 - 340	8 - 49	10	<0.2 - 34	0.8 - 4.9
Silver	<1	<1	50	<0.02	<0.02

[a] Reference 7 gives these standards "...for use in determining whether solid waste disposal activities comply with ground water criteria." Standards included in Reference 7, but not measured in these tests, are for fluoride: 1400-2400 μg/l (depending on temperature), and for nitrate (as N): 10,000 μg/l.

[b] Reference 8 contains an amendment to the chromium criteria for the EP, revising it from total chromium to Cr(VI); since the total chromium values were measured by atomic absorption (AA), the measured ranges represent upper limits for Cr(VI) in the samples.

Allen - The results indicate the following:

1. Leachate generated within the ash ponds contain elevated (over background) concentrations of several waste-related chemical constituents (e.g., boron, sulfate, calcium, arsenic). However, the surrounding soils have attenuated significant fractions of leachate constituent contaminants within the immediate vicinity of the ponds.

2. Leachate water from the upgradient portions of the ash ponds has not moved sufficiently to create steady-state concentrations of unattenuated constituents (e.g., sulfate) in the downgradient wells. However, concentrations of these constituents are expected to only reach or barely exceed secondary drinking water standards (e.g., for sulfate, 250 ppm).

Elrama - The results indicate the following:

1. Prior to disposal of FGC wastes, much of the site was contaminated by acid mine drainage, resulting in low pH (4.5 to 5) and high concentrations of chemical constituents (e.g., approx. 2000 ppm sulfate) in the ground water.

2. The landfill and runoff collection ponds may serve as potential sources of some constituents via leachate and overflow, including chloride and calcium (both at approx. 500 ppm). However, neither chloride nor calcium is at a level to cause major concern. In addition, an elevated level (approx. 0.2 ppm) of arsenic was measured at one waste/soil interface lysimeter; however, it does not appear to be a general problem. In any event, substantial attenuation of arsenic by soils at the site is expected.

3. The relative absence of elevated levels of these waste-related constituents in downgradient ground water may be explained by the relatively short time the fill has been in operation (4 years), chemical/physical attenuation phenomena (including the effects of the treatment/disposal process), or a combination of these factors.

4. The landfill does not appear to alter significantly the local concentrations of some constituents (such as sulfate) potentially available from both mine drainage and FGC wastes.

Dave Johnston - The results indicate the following:

1. The water balance and estimates of plume arrival time indicate that the widespread measurement at the site of what might elsewhere be considered elevated chemical constituent levels (e.g., sulfate, approx. 1000 ppm) is not due to the waste landfills. The estimates of plume arrival time for the

peripheral wells downgradient from (not directly under) the active landfill are in excess of 100 to 300 years considering only travel time in the saturated zone. Travel time from the 20-year old inactive landfill to a much closer (to the landfill) downgradient well is estimated to be in excess of 20 years, accounting for both unsaturated and saturated zone travel.

2. Most of the "elevated" measurements reflect pervasively high background levels characteristic of highly mineralized ground water in many western settings. However, lower measured values (e.g., sulfate, approx. 100 ppm) at one background and one peripheral well indicate that even in highly mineralized arid areas there may be areas of good water quality that require protection in waste disposal site planning and management.

Sherco - The results indicate the following:

1. Leachate movement from the ponds has thus far been sufficiently retarded by the clay liner to preclude development of significant elevations of chemical constituents in the leachate at downgradient wells.

2. There is a waste-related influence reflected in the slightly elevated levels (of boron and sulfate) measured in the peripheral/downgradient wells to the west and southwest, but it is not clear whether this is due to past leakage from the sheet piling/conduit area, to leachate that has moved through the liner, or to a combination of these two sources.

3. Because of the pervious soils in the area of the site, significant increases in concentrations of major soluble species are expected to occur in downgradient wells in the next few years. Secondary drinking water standards are expected to be exceeded in these wells. However, any effects of movement of these species off-site will be mitigated (diluted) by the Mississippi River, which flows by the plant.

4. The higher concentrations of waste parameters in FGC pond supernatant versus underlying waste interstitial waters may be due to two factors: first, the conversion by the utility to a system involving recycle of the FGC waste transport water would have resulted in increased concentrations of chemicals in the water; and second, the evaporation of water in the pond would also increase remaining chemical concentrations.

Powerton - The results indicate the following:

1. Although the completed landfill was supposed to have a 0.25 m (8 in.) Poz-O-Tec liner, during the coring operation a general absence of liner material was observed. This observation is consistent with the practical difficulty of achieving uniform placement of such a relatively thin layer of soil-like material over a large area. Current engineering practice suggests that a minimum thickness of 0.45 to 0.60 m (18 to 24 in.) of liner placement would be desired to ensure full effectiveness.

2. The surface water analytical results for Lost Creek are consistent with the water balance calculations. Both sets of results indicate that the stream has adequate assimilative/dilution capacity to lower current concentrations of chemical constituents in leachate reaching Lost Creek to insignificant levels.

3. The results also suggest that the stream, if an effective ground water flow divide, may limit the extent of further downgradient ground water contamination by the waste plume.

4. The general lack of elevated trace metal concentrations in ground water suggests that a combination of chemical attenuation (especially for chromium and lead) and dilution is preventing the release of significant quantities of these elements and/or elevation to significant concentrations at downgradient locations.

5. Elevated concentrations of nitrate at various sampling locations at the site can be attributed to local agricultural and urban nonpoint source activities and not the coal ash landfill.

Smith - The results indicate the following:

1. There appears to have been a steady state achieved between the concentrations of soluble species in the pond and in the immediately adjacent downgradient areas.

2. There appears to be little nor no chemical attenuation of the major tracer species such as calcium and strontium, but rather a progressive reduction in concentrations in the downgradient direction. This is consistent with what would be expected due to admixing of leachate with the greater amounts of dilution water.

3. The use of high total dissolved solids Bay water in the pond for makeup and its presence in adjacent downgradient areas create a situation where little incremental effect is detectable from such typical ash pond "tracer" species as sulfates, chlorides, and boron.

GENERIC ENVIRONMENTAL EVALUATION OF COAL ASH AND FGD WASTE DISPOSAL

The environmental effects of solid waste disposal practice is determined by three factors: waste type, disposal method, and the environmental setting. The data base from this project and other related projects suggests that present and future practices of coal ash and FGD waste disposal may be effectively evaluated through a matrix consisting of four waste types, three disposal methods, and five environmental settings.

The four waste types are:

1. Fly ash or fly ash admixed with other materials. A significant body of literature suggests that the majority of trace metals available for leaching from utility solid wastes may be associated with those containing fly ash. Thus this category of wastes includes fly ash or fly ash mixed with bottom ash and fly ash/bottom ash/FGD waste mixtures (excluding chemically treated FGD wastes; see item 3, below).

2. Non fly ash materials. In this category are included bottom ash (or boiler slag) and FGD wastes that are disposed of separately from fly ash (including forced oxidation wastes). This category usually contains lesser concentrations of trace metals, but can result in higher concentrations of major species (e.g., chlorides from FGD waste).

3. Stabilized FGD wastes. FGD wastes may be processed or stabilized for full-scale disposal by a variety of processes; the processes presently in commercial practice involve the addition of lime and fly ash, or processed slag. Lime/fly ash stabilization for landfill disposal is presently practiced at some power plants and is expected to grow in importance. Processed FGD wastes are a separate category because of the differences in their physical and chemical properties created by the stabilization process.

4. Dry FGD wastes. Several dry FGD systems are expected to come into commercial use over the next 3 years.(2) It appears that calcium-based dry FGD systems are anticipated to grow more than the sodium-based systems. By either process, dry FGD systems provide a combined waste containing fly ash and the sulfur compound in a relatively dry form that is likely to be sent for disposal to a managed landfill. The physical and chemical properties of these wastes are expected to be different from the other categories discussed above; additionally, there is a relative lack of even limited field scale information on their leaching characteristics to date.

There are three disposal methods for coal ash and FGD wastes that are in practice and expected to continue in the future: (1) pond disposal; (2) interim ponding followed by landfill disposal; and (3) landfill disposal (including disposal in mines, which is considered a special case of landfilling).

Three of the five environmental settings for solid waste disposal are based on major differences in climate and hydrogeology. These are: (1) coastal areas, specifically those areas where surface water and ground water are influenced by the ebb and flow of tides; (2) arid areas, characteristic of much of the western U.S. where net evaporation generally exceeds precipitation by a significant margin; and (3) interior areas, characteristic of the non-coastal portions of the eastern U.S. where there tends to be more of a balance between precipitation and evaporation and where permanent surface water bodies are in such abundance as to be near many disposal sites.

Further evaluations during this project suggested that a further breakdown of two special categories would be useful because of their significant characteristics. These are: (1) arid areas in the west where ground waters and surface waters are very highly mineralized, and (2) interior areas subject to acid mine drainage. Both of these last two types of settings and the coastal setting tend to have water quality characteristics that can potentially show less of an incremental effect from coal ash and FGD waste leachates. This is because the waters in these areas already contain a number of chemicals found in the leachate.

Table 7 is a matrix of waste types, methods of disposal, and environmental settings and indicates combinations for which field-scale and other information is available. Sources of data and information other than this study included the Utilities Solid Waste Activities Group (USWAG), the Electric Power Research Institute (EPRI), and the Department of Energy (DOE). DOE is currently sponsoring a study of disposal FGD wastes in a surface mine; the study was originally sponsored by EPA. As is clear, some information is available for most of the combinations that are being practiced today or are likely to be practiced in the future.

It appears that, on balance, technology exists for environmentally sound disposal of coal ash and FGD wastes using any of the modes of disposal. Potential environmental effects are highly site and system specific. For some combinations of waste types, disposal methods, and environmental settings, mitigative measures must be taken to avoid ground water and/or surface water contamination. However, site specific application of good engineering design and practice can mitigate most potentially adverse environmental effects of waste disposal.

TABLE 7. SUMMARY OF INFORMATION AVAILABLE FOR COMBINATIONS OF WASTE TYPES, DISPOSAL METHODS, AND ENVIRONMENTAL SETTINGS

	Ponding				Interim Ponding/Landfilling				Landfilling			
	Fly Ash^{a}	Non-Fly Ash^{b}	Processed FGD	Dry FGD	Fly Ash^{a}	Non-Fly Ash^{b}	Processed FGD	Dry FGD	Fly Ash^{b}	Non-Fly Ash^{b}	Processed FGD	Dry FGD
COASTAL SETTING	X Smith^{c}	P	NA	NA	X/P^{d} Chisman Cr. (USWAG)	P	P	NA	P	P	P	P^{e}
ARID WESTERN SETTING- Not Highly Mineralized	P	P	NA	NA	P	P	P	NA	P	P	P	P^{e}
ARID WESTERN SETTING - Highly Mineralized	P	P	NA	NA	P	P	P	NA	X Dave Johnston; Milton Young (DOE/EPA)	P	P	P^{e}
INTERIOR SETTING - Not Highly Acidic	X Allen, Sherco, Michigan City (USWAG), Wallingford (USWAG)	P	X Bruce Mansfield	NA	X/P^{d} Bailly (USWAG)	P	P	NA	X Powerton, Zuellinger (USWAG) Hunts Brook (USWAG) Dunkirk (DOE)	P	X Conesville (EPRI/USWAG)	P^{e}
INTERIOR SETTING - Highly Acidic (mine drainage)	P	P	P	NA	P	P	P	NA	P	P	X Elrama	P^{d}

Notes:
a. Includes co-disposal of fly ash with other wastes.
b. Includes FGD wastes without fly ash and bottom ash.
c. Plants for which data and information are obtained are listed in their appropriate positions.
d. Either the interim pond or landfill aspect of operation studied at field scale, but not both.
e. Laboratory data only.

Key:
X = Data available from full-scale field studies.
P = Data available from laboratory and/or limited-scale field studies for projection purposes.
NA = Matrix combination not applicable due to lack of present and future practice.

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ENGINEERING/COST EVALUATIONS

The first major efforts in the engineering/cost evaluations involved development of site-specific conceptual engineering designs and costs (capital and first year operating and maintenance costs) for the current solid waste handling and disposal operations at the six study sites. To facilitate the ultimate use of the cost data, the estimates were developed by breaking down the waste handling and disposal operations into five modules: (1) raw material handling and storage; (2) waste processing and handling; (3) waste storage; (4) waste transport; and (5) waste placement and disposal (including site monitoring and reclamation).

Based on the site-specific cost estimates and other studies by TVA, EPRI, and other organizations, generic capital and O&M cost estimates were then prepared for individual modules comprising waste handling and disposal for coal ash and FGD wastes. Tables 8 and 9 provide a summary of the results of this effort.

The range of costs given represents variations in specific plant operations as well as variations in the several cost estimates used in preparing these estimates. For example, the higher end of the range for FGD waste handling/processing might include thickening, vacuum filtration, and mixing with lime and fly ash, while the lower end could represent a simpler operation with little or no processing. Figures 2 and 3 show, in graphical form, the estimates for the FGD waste handling/processing "module." Similar figures for all the modules listed in Tables 8 and 9 will be included in the final report for the project.(4)

CONCLUDING REMARKS

Results from this 3-year study of disposal of coal ash and FGD wastes from coal-fired electric generating plants should provide major technical guidance for regulatory bodies and the utility industry. However, results from field studies of this type are limited, and predictive tools (e.g., computer models) for evaluating interactions between these wastes and site-specific hydrogeologic systems are, in many cases, inadequate. For this reason, additional efforts sponsored by the industry are currently underway to develop more sophisticated tools for predicting and analyzing the potential environmental effects of coal ash and FGD waste disposal. These efforts will be described later in this session.(9)

TABLE 8. GENERIC CAPITAL COST ESTIMATES FOR FGC WASTE DISPOSAL
(Late 1982 Dollars)[a]

		Capital Cost Range ($/kW)			
		Plant Size (MW)			
Module	Submodule	250[b]	500[b]	1000[b]	2000[b]
Fly ash handling/processing	Wet handling w/o recycle	2.3-4.3	1.9-3.5	1.5-2.9	1.3-2.3
	Wet handling w/recycle	3.7-6.8	3.0-5.5	2.4-6.4	1.9-3.6
	Dry handling	2.2-4.1	1.8-3.3	1.4-2.7	1.2-2.2
Fly ash storage	Dry	4.7-8.8	4.2-7.7	3.7-6.8	3.2-5.9
Fly ash transport	Wet sluicing	3.5-6.4	2.7-5.1	2.2-4.0	1.7-3.2
	Dry trucking	0.3-0.5	0.3-0.6	0.3-0.5	0.2-0.5
Fly ash placement/disposal	Unlined pond	15.1-27.8	12.9-23.9	11.0-20.5	9.4-17.5
	Landfill	4.3-8.1	3.3-6.1	2.5-4.7	1.9-3.6
Bottom ash handling/processing	Wet handling w/o recycle	2.2-4.1	1.7-3.2	1.3-2.5	1.0-1.9
	Wet handling w/recycle	2.5-4.6	2.0-3.7	1.6-3.0	1.3-2.4
Bottom ash transport	Wet sluicing	3.0-5.6	2.4-4.5	1.9-3.6	1.5-2.8
	Dry trucking	0.2-0.4	0.2-0.3	0.1-0.2	0.1-0.2
Bottom ash placement/disposal	Unlined pond	6.4-11.8	5.1-9.6	4.2-7.7	3.4-6.2
	Landfill	1.3-2.4	1.1-2.0	0.9-1.6	0.7-1.3
Raw materials handling/storage	Dry (lime and fly ash)	2.4-4.5	2.1-3.9	1.9-3.4	1.6-3.0
FGD waste handling/processing[c]	Wet handling	18.1-33.6	15.2-28.3	12.8-23.8	10.8-20.0
FGD waste transport	Wet sluicing	0.7-1.3	0.5-1.0	0.4-0.8	0.4-0.7
	Dry trucking	0.4-0.7	0.3-0.6	0.3-0.5	0.3-0.5
FGD waste placement/disposal	Unlined pond	10.0-18.	8.9-16.6	7.9-14.7	7.0-13.1
	Landfill	4.1-7.6	3.3-6.2	2.7-5.0	2.2-4.0

[a] Engineering News Record (ENR) Index = 3931.11 (1913 - 100)
= 365.97 (1967 - 100)

[b] Relationship between plant size and waste generation for typical case:

	Annual Waste Generation Rate (dry metric tons/MW of Plant Generating Capacity)
Fly Ash	280
Bottom Ash	70
FGD Waste	240

"Typical Case" Assumptions

Coal Properties:	2% S, 13% Ash, 10,300 Btu/lb (24.4 M MJ/kg)
Load Factor:	70%
Heat Rate:	10,250 Btu/kWh (10.8M MJ/kWh)
SO_2 Removal:	90%
Lime Stoichiometry:	1.1
Fly Ash/Bottom Ash Ratio:	80/20

[c] Assumed FGD System: Wet Lime Scrubbing

TABLE 9. GENERIC ANNUAL COST ESTIMATES FOR FGC WASTE DISPOSAL
(Late 1982 Dollars)[a]

		Annual Cost Range ($/dry metric ton)			
		Plant Size (MW)			
Module	Submodule	250[b]	500[b]	1000[b]	2000[b]
Fly ash handling/processing	Wet handling w/o recycle	2.5-4.6	1.0-3.7	1.6-3.0	1.3-2.3
	Wet handling w/recycle	3.7-6.8	2.9-5.4	2.3-4.3	1.8-3.6
	Dry handling	2.5-4.7	2.1-3.9	1.7-3.2	1.5-2.7
Fly ash storage	Dry	3.3-6.1	3.0-5.6	2.8-5.2	2.5-4.7
Fly ash transport	Wet sluicing	4.2-7.6	3.2-5.9	2.5-4.7	2.0-3.7
	Dry trucking	1.7-3.1	1.5-2.8	1.3-2.5	1.2-2.2
Fly ash placement/disposal	Unlined pond	11.5-21.3	9.1-16.8	7.2-13.5	5.7-10.5
	Landfill	7.0-13.0	5.6-10.5	4.6-8.5	3.7-6.9
Bottom ash handling/processing	Wet handling w/o recycle	11.3-21.0	9.0-16.7	6.9-12.8	5.3-9.9
	Wet handling w/recycle	12.3-22.8	10.3-19.1	8.4-15.7	6.9-12.8
Bottom ash transport	Wet sluicing	9.2-17.1	7.3-13.5	5.6-10.3	4.3-7.9
	Dry trucking	3.4-6.3	2.8-5.2	2.2-4.1	1.8-3.3
Bottom ash placement/disposal	Unlined pond	9.2-17.1	7.9-14.6	6.5-12.1	5.4-10.0
	Landfill	5.4-10.0	4.7-8.8	4.1-7.6	3.5-6.5
Raw materials handling/storage	Dry (lime and fly ash)	4.1-7.6	3.7-6.7	3.4-6.2	3.0-5.6
FGD waste handling/processing[c]	Wet handling	17.2-31.9	13.8-25.5	11.0-20.5	8.8-16.4
FGD waste transport	Wet sluicing	1.1-2.1	0.9-1.7	0.7-1.3	0.6-1.1
	Dry trucking	2.9-5.4	2.3-4.3	1.8-3.3	1.4-2.6
FGD waste placement/disposal	Unlined pond	8.5-15.8	6.7-12.4	5.2-9.7	4.1-7.6
	Landfill	4.0-7.5	3.4-6.3	2.8-5.3	2.4-4.4

[a] Engineering News Record (ENR) Index = 3931.11 (1913 - 100)
= 365.97 (1967 - 100)

[b] Relationship between plant size and waste generation for typical case:

	Annual Waste Generation Rate (dry metric tons/MW of Plant Generating Capacity)
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FGD Waste	240

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Load Factor:	70%
Heat Rate:	10,250 Btu/kWh (10.8M MJ/kWh)
SO_2 Removal:	90%
Lime Stoichiometry:	1.1
Fly Ash/Bottom Ash Ratio:	80/20

[c] Assumed FGD System: Wet Lime Scrubbing

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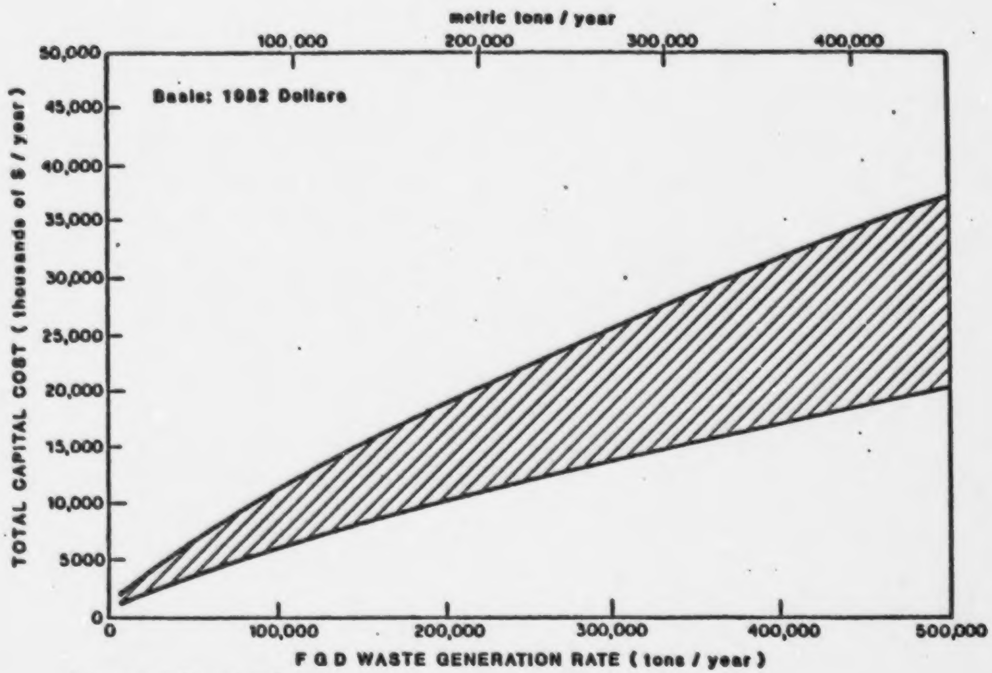
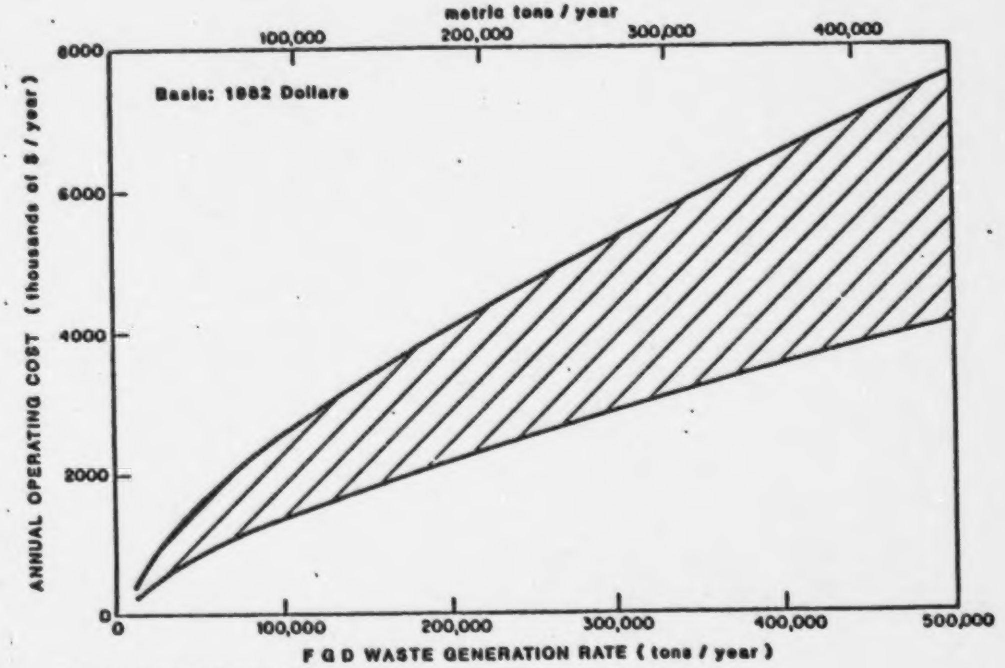


Source: Arthur D. Little, Inc. Estimates

FIGURE 2. FGD WASTE HANDLING AND PROCESSING: ANNALIZED COSTS VERSUS FGD WASTE GENERATION RATE



* Source: Arthur D. Little, Inc. Estimates

FIGURE 3. FGD WASTE HANDLING AND PROCESSING: ANNALIZED COSTS VERSUS FGD WASTE GENERATION RATE

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OPERATIONS HISTORY OF LOUISVILLE GAS & ELECTRIC FGD SLUDGE STABILIZATION

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ABSTRACT

The Louisville Gas & Electric Company (LG&E) has been an industry leader in the deployment of flue gas desulfurization technology from its very inception. LG&E was one of the first major utilities to install SO_2 scrubbers; and, at the present time, it operates seven scrubbers serving a combined generating capacity exceeding 2200 m.w. These scrubbers which have been purchased from several different manufacturers, represent a broad spectrum of scrubber technologies including dual alkali, lime and limestone systems.

The bleed stream from these scrubbers is dewatered, and the resulting solids are chemically and physically stabilized in processing plants purchased from Conversion Systems, Inc. (CSI), an early pioneer in the field. Unit #6 at Cane Run, a 270 m.w. unit with a dual alkali scrubber, is served by a stabilization facility which began operations in April, 1980. Units #4/5 at Cane Run have a combined capacity of 360 m.w. and are equipped with lime scrubbers. The waste from these scrubbers is combined and treated in another stabilization facility. The four units at Mill Creek have a combined capacity exceeding 1600 m.w. and are also equipped with lime scrubbers. One large stabilization facility was installed to handle the combined bleed from all four Mill Creek scrubbers.

The stabilization plants have run for approximately three years and are operated and maintained by LG&E personnel with advisory assistance supplied by CSI. The knowledge gained from the experiences of LG&E and CSI in the last three years of operation would be useful both to current operators of stabilization facilities and to those who are anticipating the procurement of scrubbers and stabilization facilities.

This paper presents some of the operating and maintenance history which has been obtained from these plants. Included will be discussions on reliability of individual equipment and discussions of some modifications which were

made to improve reliability. Operating problems will be addressed including some of the inherent difficulties encountered in processing scrubber sludge and fly ash.

In addition, a short discussion of the landfill operation which is an integral part of the stabilization process will be presented, as well as the results of several years of landfill investigation. Discussions of the impact of the landfill operation and plant control upon the environmental properties of the final landfilled material will be offered.

BACKGROUND

Louisville Gas & Electric Company (LG&E) owns and operates two coal fired generating stations in Jefferson County, Kentucky. These stations serve the city of Louisville and its surrounding suburbs. During the early 1970's, when sulfur dioxide emissions from coal burning power plants became a high environmental priority, LG&E was required to install scrubbers on the newest of its coal fired generating units. At that time there had been only a handful of scrubbers installed on major utility power plants. LG&E became one of the industry leaders with its early demonstration scrubber at their Paddy's Run Plant. It has continued in the lead position by installing a total of seven additional scrubbers at its Cane Run and Mill Creek Generating Stations.

During the early operating days of the Cane Run scrubbers, it became apparent that some form of sludge stabilization was necessary to dispose of the scrubber bleed and to prevent the creation of a surface or groundwater pollution problem. It was for this reason that LG&E in the late 70's contracted with Conversion Systems, Inc. (then known as IU Conversion Systems) to design, engineer and assist in the erection of three scrubber sludge dewatering and stabilization facilities. Two are located at Cane Run; and, one at Mill Creek. These three stabilization facilities are operated by manpower provided by LG&E with the help of an advisory team provided by CSI. The disposal of the scrubber waste is accomplished by dewatering and stabilization, utilizing the Poz-O-Tec system developed by CSI together with controlled landfill disposal.

The LG&E/CSI operations team has acquired over three years experience in operating these three stabilization facilities and has encountered many operation and design problems. Many of these problems will be encountered by other firms as they begin operation of their own stabilization plants.

The first of these facilities was a relatively low budget blending unit designed to handle waste generated by the Cane Run #6 unit. This unit is a 270 megawatt generator using coal with 12 to 16% ash and 4% sulfur content. It is equipped with an electrostatic precipitator and one of the first dual alkali scrubbers. The blending facility included fly ash and lime storage and the feed systems necessary to blend fly ash and lime with dewatered sludge from the scrubber. The interface on this plant was somewhat unusual when compared to other facilities provided by Conversion Systems. LG&E had been dewatering thickener underflow and disposing of this material in their bottom ash ponds with no stabilization. When it was decided to install stabilization, a facility was designed which placed the blending equipment within the skirt of the fly ash silo. This compact arrangement, although economical and expedient, was later the cause of many operational and maintenance headaches. (See Figure 1)

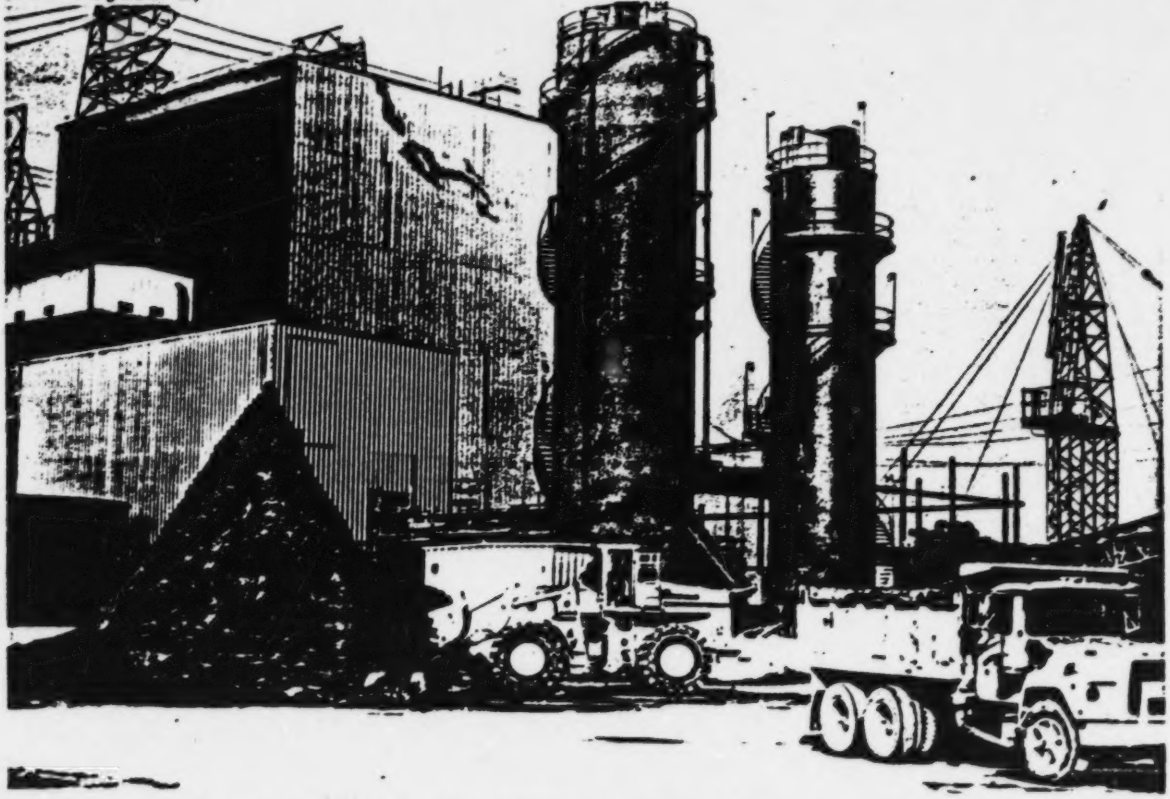

FIGURE 1 - Exterior of Cane Run #6 showing Poz-O-Tec being loaded for transport to landfill

UNDERFLOW HANDLING

In order to properly evaluate the experiences encountered in the operation of the Cane Run facility, it is necessary to consider the system including not only the blending unit, but also the vacuum filters and thickeners. Those familiar with FGD thickener operation realize that it is difficult to maintain a specific underflow density in normal operation. Usually the variance in underflow density is alleviated by a surge tank between the thickener and the vacuum filters in which the thickener underflow is blended to control the density and flow. In the Cane Run #6 facility the vacuum filters are fed directly from the thickener underflow pumps. Since the thickener underflow will vary from 5% to 30% solids, the resulting filtercake solids content is also highly variable. Because the processing plant must handle upsets in the thickener operation, the quantity of material is highly variable. Due to wide swings in flow and solids content, the control system is sometimes unable to respond, and the final blended product is not always consistent. At times the material is very dry which causes stockpile and landfill dusting problems; and, at other times, the material is wet and sloppy which causes difficulties in handling and can turn the landfill into a quagmire. We have been able to mitigate the effects of this variability by careful stockpile and landfill management techniques. These techniques include segregation and selective blending of stockpiled material with different properties. These practices enable us to produce a more consistent landfill material which minimizes handling and dusting problems, thus increasing landfill operating efficiency.

The absence of a surge tank has increased problems in the day-to-day manpower allocation and management of the facility. The waste treatment plant is a slave to the thickener operation. The plant must be ready to receive thickener underflow whenever the torque on the thickener rake is high. The facility also must sit idle or process at a minimal rate while a bed is developed in the thickener.

The production rates of the processing facility have, at times, exceeded the design rates due to high torque conditions in the thickener. In most cases we have been able to handle these additional quantities of material, although the resulting product has not always been entirely up to specification requirements, and the overload condition has put an excessive strain on the equipment. We have been able to handle this off-spec material through the use of stockpile and landfill management techniques. These techniques have enabled us to maintain environmentally safe landfill sites.

The Cane Run #4/5 and Mill Creek Processing Plants have been equipped with underflow surge tanks which have produced a more consistent product and permits greater flexibility for maintenance scheduling.

FLY ASH HANDLING

We have installed a total of four fly ash silos at the three LG&E facilities. All of these silos are designed with mass flow bin bottoms and variable speed tapered screw feeders. Although this design was initially more expensive than the tradition air slide flat bottom fly ash silo installation, there are almost no operating costs since it eliminates the need for compressed air for air slide operation. We have also eliminated the high maintenance cost usually encountered with star valve installations. This fly ash feeder and bin design, however, is not without its operating problems. The bin bottoms are designed with long, tapered sides which reduce the potential for bridging and rat-holing by improving the flow characteristics of the fly ash. The result is a relatively uniform delivery of fly ash to the tapered screw feeder.

The tapered screw removes the fly ash at a predictable and controllable rate and feeds it into the process. During the start-up of the Cane Run #6 facility, we experienced several fly ash floods; that is, the contents of the silo flowed through the feed system totally out of control. Subsequent experiments and careful review of our operating logs indicated that this condition was observed only at times when the fly ash level in the silo was very low or when ash was being transferred into the system. Since the ash is transported via pneumatic conveyor, we theorize that air was remaining entrained in the ash and while in this state the ash was extremely flowable.

Our solution to this problem was to modify our operating procedures and our equipment. Our operating procedures were modified to require the slide gate above the tapered screw feeder to be closed whenever the silo level is below the intersection of the cone and the straight side of the silo, or whenever ash is transferred into the silo. In addition, we installed instrumentation which automatically closes the gate whenever a flood condition is detected. These measures proved successful in the Cane Run #6 plant; but because of the size of the gates in the two later plants, it was decided to install a smaller gate downstream of the feeder. This gate is designed to close whenever a flood condition is detected.

FLY ASH DUST CONTROL

A problem encountered early in start-up was the operation of the mixer dust collectors. These collectors are intended to control fly ash and lime dust (and water vapor) which escapes (the equipment) to the atmosphere at the mixer. Our original design proved inadequate due to plugging of the duct work. The failure of these collectors led to high dust concentration in the air which made it uncomfortable for operators and created problems for the equipment. A revised design was installed on the Mill Creek plant which proved successful. While earlier installations attempted to maintain a negative pressure in the mixer in an effort to contain the dust within the mixer chamber, this new design concentrated its efforts outside the mixer. Observation of the mixer system revealed that the atmosphere of the mixer is extremely turbulent with the turbulent action causing dust particles to be suspended. When this atmosphere was pulled into the dust collector, the collector soon became overloaded and clogged. We soon learned that the dust particles tend to settle, if permitted, in a still atmosphere. This was accomplished with a revised dust collection hood which was designed with a long stilling chamber immediately after the mixer. The dust collector suction was moved downstream of the stilling chamber. This design provided sufficient time for dust particles to settle and become captured by the slightly moist product. Our success with this system was extremely gratifying, and efforts are now underway to install the same system at the other Louisville plants.

MAINTENANCE

The fly ash laden atmosphere combined with the corrosive properties of lime and the water in the filtercake all combine to make a maintenance headache. Those who have never operated a sludge stabilization facility should not be lulled into believing that these plants do not require operators, or that maintenance will be minimal. Improvements in the fly ash dust collection system and other equipment modifications have contributed greatly to improving the service life of our equipment. However, additional maintenance routines initiated by LG&E at these plants have been an even greater factor in the service life of this equipment. A good deal of the maintenance problems encountered in a sludge stabilization facility are due to ineffective or inadequate cleanup efforts. The theory of operation of the Poz-O-Tec process is that the fly ash and lime chemically combine to form a matrix which binds the FGD scrubber sludge into a cement-like material. This material will build up and cure into a hard, abrasive mass which will destroy bearings, gears and other mechanical parts. This was evident in the early days at Louisville when material had been allowed to cure in the mixer. The mixer paddles began to hammer the cured material. The resultant stress was transmitted along the shaft and eventually contributed to the premature failure of several gear teeth in the mixer speed reducer. LG&E has instituted a program which keeps the material build-up under check. Every day the plant is shut down for complete cleaning. This includes cleaning of the mixers and hosing of conveyor parts. (See Figure 2) This cleaning usually takes several hours but has proven worthwhile in extending the life of the equipment. To speed cleanup at the Cane Run #6 processing plant, the mixer has been coated with a high density polyurethane coating which has proved to be very useful in preventing material build-up. It permits the mixers to be cleaned with high pressure water. This coating has, however, become scored by the fly ash and its benefit will diminish. A reapplication will hopefully restore the surface. If this coating is successful, the other processing facilities will have this coating installed.

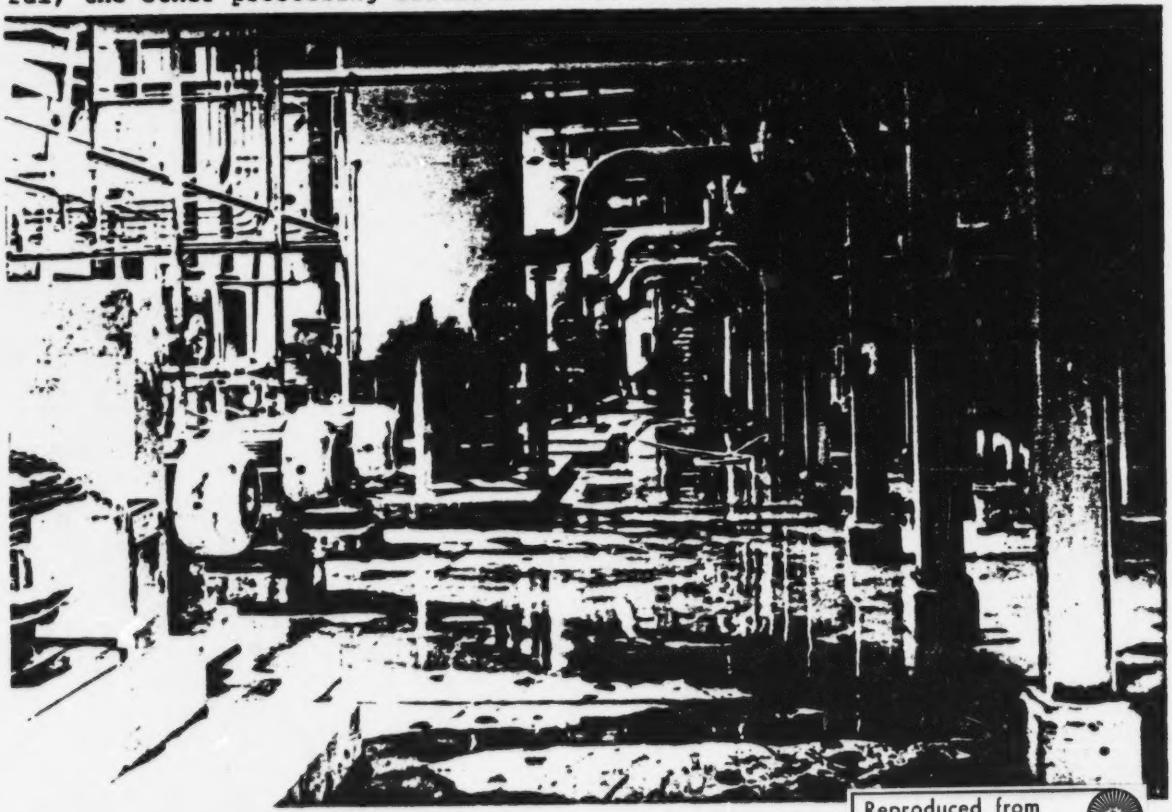

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FIGURE 2 - Interior of Mill Creek During Clean-Up

The lubrication program developed for the Louisville plants has recently been modified to take into account some of the lessons we have learned. As was stated earlier, the facilities are washed down at least once a day with water at about 50-75 psig. Over a period of time daily washdown will flush the grease from bearings. The fly ash and lime present in the atmosphere may also contaminate grease and oil. These contaminated lubricants if left in place act as abrasives, reducing bearing and gear life. To eliminate these problems, we have gone to a lubrication schedule which reduces the time between the application of grease and increases the frequency of gearbox oil changes.

WINTER OPERATION

Those stabilization plants operating in cold climates are faced with the additional hardship of winter operation. Cold weather hampers processing plant operation by reducing the filterability of the thickener underflow. In addition, the problem of belt and idler freezing reduces conveyor belt and idler life. The clean-up of outdoor conveyors is also a problem since the use of high pressure water hoses increases the possibility of ice-up.

Cold weather further inhibits the fly ash/lime reaction. This particular problem is resolved by stockpile management techniques which utilize the product's insulating properties. With proper techniques, even if the outside temperature is below freezing, temperatures approaching 100°F may be maintained within the pile. Periodic monitoring of these temperatures, however, is necessary in order to insure sufficient reaction before landfilling the product and to guard against over-reaction which would result in the creation of boulders.

SOLUTIONS

In the design and construction of the processing plant that services the Cane Run Unit #4/5, several modifications were made and improvements realized over the Cane Run #6 design. This unit is about 3/4 mile from the thickeners and approximately 1 mile from the fly ash silos. It was decided to include 100% redundancy for the filtration system, and a single surge tank was installed. No redundant conveyor systems or redundant mixers were installed. Operation of the surge tank has enabled LG&E to schedule operations so that most of the maintenance is performed during the daylight shift. It has also smoothed the flow variations which have been seen in the Cane Run #6 unit. Underflow feed from the surge tank varies approximately 2 to 3% over the duration of a pumping shift. We are able to process at a more consistent rate and generate product with a more consistent solids content.

The processing plant servicing the Mill Creek Units #1, 2, 3 & 4 was designed with almost 100% redundancy. The only common element at this plant is the large surge tank. Redundant fly ash and lime systems are included in this facility as well as separate conveyors and mixer trains. Separate mixers allowed the plant to schedule routine cleaning and maintenance. This insures efficient cleaning of each mixer since the cleaning crews are able to take their time and do a complete job.

The Mill Creek facility is also equipped with an on-site laboratory which is manned during the daylight shift. This provides immediate feedback to the processing plant operator so that variances in design mix are discovered quickly and are corrected. Because of this LG&E is assured of efficient utilization of lime and material with the proper in-place qualities.

OTHER RECOMMENDATIONS

The utility just beginning to operate its first FGD sludge processing facility should realize that many elements will effect operating efficiency. At LG&E several types of filtercloth were utilized based upon filter-leaf tests performed by CSI's on-site advisors. The optimum cloth was chosen based upon filtercake production rates and properties. At Mill Creek we have utilized several types of filtercloth on different filters. Each type is used for different properties of the thickener underflow. We have done this because we understand that there are variabilities in the coal burned and the operation of the scrubber which must be taken care of by the sludge stabilization facility. Since all material must be processed by the sludge stabilization facility, this facility must be adaptable to the variances which will be encountered in the operation of the power plant, and it must be ready to accept upsets with- out causing plant downtime. Therefore, we have tried to adapt the plants to accept these variances. There are, of course, some variances which go beyond the adaptability of the processing facility. We have been able to handle these through the use of stockpiling and landfilling techniques, most notably microencapsulation, which we have successfully employed at the Louisville facilities.

LANDFILL

The above topics discussed the processing facilities required to process the FGD scrubber sludge into a stabilized material. However, to complete properly the Poz-O-Tec system, the processed material must be conditioned and placed within a permitted landfill. LG&E also contracted with Conversion Systems to provide a Waste Management Plan for the disposal of the material produced by their processing facilities. This plan included all necessary information and procedures for LG&E to obtain landfill permits from the State of Kentucky for both of their generating stations. The plan included proposed material characteristics (unconfined comprehensive strengths, permeability, leachate), landfill and process plant quality control programs, and landfill site development plans. With this information, LG&E was able to obain landfill permits at both power stations.

At this time, landfills are being developed at both sites in accordance with the Waste Management Plan with monthly inspections by the State of Kentucky. The permitted sites include borrow areas (areas where soil was removed to construct the station), flood plain areas of the Ohio River and other areas not suitable for the construction of power plant structure. One of the areas not suitable for utility construction is currently being reclaimed by landfilling with Poz-O-Tec at the Mill Creek generating station. Plans have been made to construct a limestone crushing plant upon completion of this landfill site.

GROUND AND SURFACE WATER MONITORING

Included in the Waste Management Plan were programs for ground and surface water monitoring. The groundwater monitoring is accomplished by quarterly sampling of wells located near the landfill sites. These locations were established by the groundwater flow which flows toward the Ohio River. Efforts were made to locate one monitoring well upstream of each landfill site and two or three wells downstream of each landfill site in order to determine the effect of the landfill upon groundwater quality. Initially, the wells were sampled to obtain a composite standard against which subsequent water quality data could be measured. Once the landfill operation started at each site, the up-gradient wells have been used to monitor any changes in the quality of groundwater passing beneath the disposal area. Before each sampling, the static water level in each monitoring well has been recorded. This data has been used to determine the annual fluctuations in the groundwater elevation for the disposal area, since the placement of processed material within the disposal area should not affect the existing dominant direction of the groundwater flow. Up to this time the water quality analysis have indicated no change in the overall groundwater quality and dominant direction of the groundwater flow.

The obtaining of the monitoring well samples has been simplified as each well has had a submersible pump installed. This pump is energized by a portable generator which enables the laboratory technicians to sample each well in less than one hour. Originally, the wells were sampled with a siphon pump located on the back of a pick-up truck. Priming of each well posed problems due to the wells isolated location which increased the time required to sample each well. After obtaining the samples, they are preserved and sent to the CSI's Technical Center for analysis in accordance with the permit guidelines.

All landfill run-off caused by precipitation is controlled and routed through a pond constructed near each landfill site. The ponds were designed to collect the run-off and discharge through a sandfilter. Sizing of the ponds and associated draining channels were in accordance with the Louisville area one hundred year rainfall criteria. Development of the landfill incorporated drainage designs to control all run-off during construction. Efforts have been made daily to control sediment due to errosion as an excessive amount of sediment in the run-off increases the maintenance of the pond and sandfilter. Up to this point, the sandfilter has had to be cleaned approximately twice a year. Run-off samples have been obtained at weirs installed at each sandfilter discharge, after each rainfall.

QUALITY CONTROL

In order to continue to meet the parameters of the landfill permits, the Waste Management Plan also provides a quality control program for each processing facility and landfill site. These programs are maintained by LG&E's laboratory personnel with CSI's Advisory Staff personnel providing any technical advice needed to maintain these programs.

Daily process plant quality control activities include sampling of all the raw materials sent to the process plant and determining the proper addition rates for the process plant operators to utilize. The operators use these recommendations to make minor set-point changes on automatic controllers. At times, major set-point changes are required when one of the raw materials is received out of specification. This usually occurs when the filtercake solids are low due to low solids received from the thickeners. Other daily routines include sampling of the stockpiled material to determine its curing rate. This analysis also provides information on back-shift production since the laboratory personnel are only assigned to the process plant during the daylight shift. This minimum coverage of quality control is satisfactory as long as the process plant personnel operators operate the plant in the automatic mode.

Whenever non-specification material is produced due to operator error, equipment breakdowns, or raw material availability, the quality control personnel recommend set-point changes to produce an above specification material to blend with the below specification material. Cooperation between both the operators and quality control technicians has resolved most deviations smoothly. (See Figure 3)

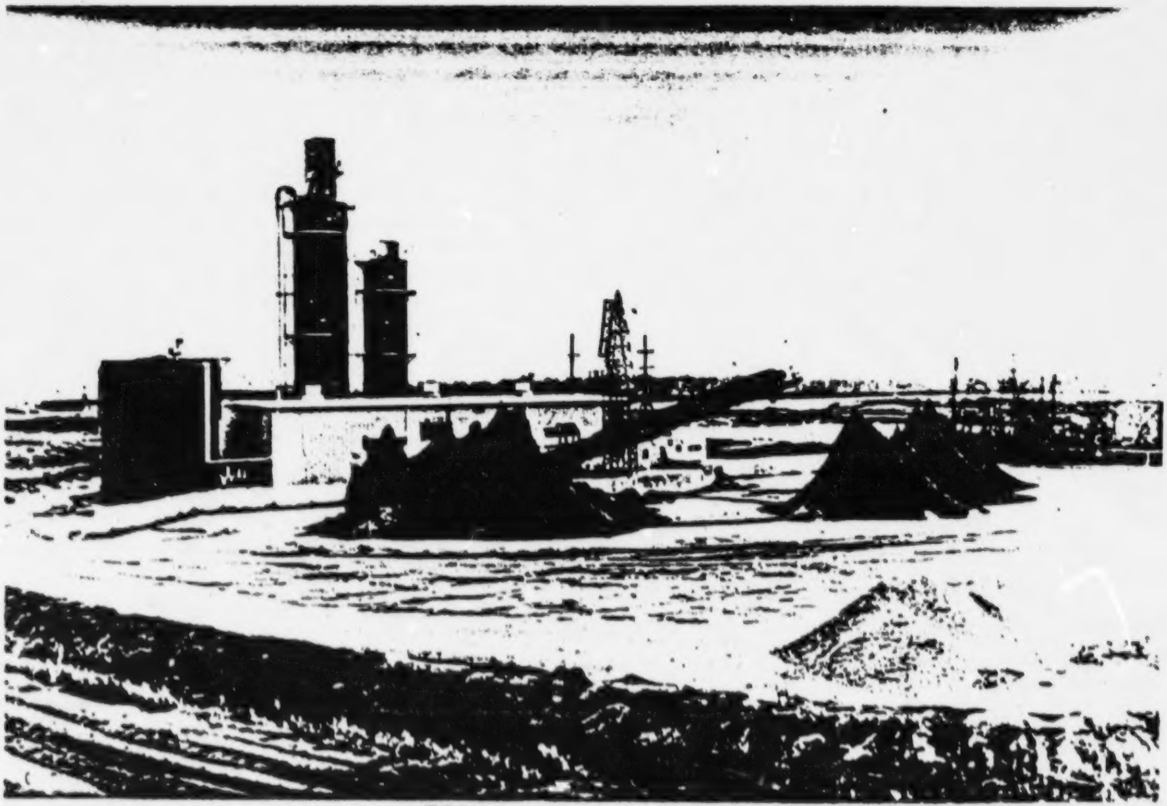

FIGURE 3
Exterior of Cane Run #4/5 Showing Stockpile Management Practices

During the start-up of each facility, ratio tests were performed to determine minimum addition rates of fly ash and lime. It was found that for these particular installations, the fly ash addition rate could be reduced to 60% of the design with a small increase in the addition of lime and still meet the minimum material characteristics required in the landfill.

The landfill quality control program is also maintained by LG&E personnel with assistance from the CSI Advisory Staff personnel. The same technicians that perform the process plant analysis also maintain the landfill program since the frequency of testing is dependent upon the amount of processed material produced. In-place landfill density tests are performed for every 10,000 to 12,000 tons of material produced. These tests are used to control the landfill placement technique and to insure that the material is placed and compacted to a minimum density specification. When this minimum density specification has been achieved, it has been found the landfill strength permeability and leachate criteria are met. These three criteria are continually checked as landfill test cylinders are prepared routinely for testing at the CSI's Technical Center. These test cylinders are molded to the landfill in-place density test results by the LG&E technicians and, after curing, are sent to the Technical Center for analysis. Results from these landfill test cylinders have shown that as long as the processed material of the proper mix is properly conditioned, placed and compacted, all permit guidelines concerning the materials physical characteristics have been met.

LANDFILL OPERATIONS

LG&E contracted a local heavy-equipment contractor to prepare the landfill sites and place the processed material in accordance with the Waste Management Plan. Since the landfills are located on the utility property next to the generating station, the contractor has been able to utilize offthe-road equipment. The size of his fleet of hauling equipment has grown as each of the three processing facilities started up. At this time, he is using both 35 yard and 50 yard capacity trucks. (See Figure 4) The larger trucks have posed a problem as the roads bearing this traffic must be well-constructed to support their gross weight. In the landfill itself, additional effort must be made in the site development to eliminate truck traffic on recently placed material as it will not initially support this weight. Since curing rates are slowed during winter operations, the active landfill must be increased accordingly.

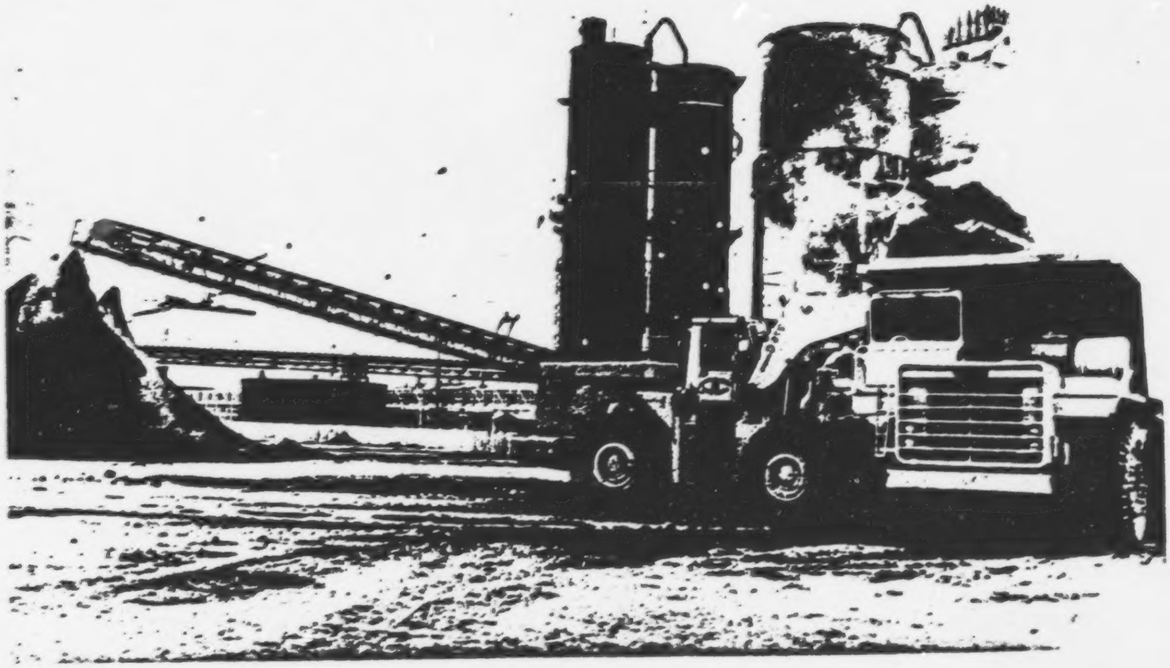

FIGURE 4 - Exterior of Mill Creek showing Poz-O-Tec being loaded for transport to Landfill

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Low-ground pressure bulldozers place and compact the material with a self-propelled vibratory roller providing additional compaction and surface sealing. Maintenance repairs on the landfill equipment is actually less than the contractor's normal equipment repairs on earth projects. This can be attributed to the fact that although the processed material has abrasive additives, it is less abrasive than the sandy soil in which his equipment usually operates. The contractor operates five days a week, approximately eight hours a day. Depending on which site he is actively working, his hauling rate varies between 200 to 500 tons per hour.

MATERIAL CHARACTERISTICS

The LG&E material characteristics are based upon the type of coal used, type of scrubber and agent, fly ash collector size and collection efficiency and available lime supply. The design specifications of process plant mixes and ultimate landfill characteristics were achieved without any major changes from the original Conversion Systems, Inc. equipment design. We did, however, lower the filtercake solids content and production range, as it was found during start-up at all three plants that the filtercake could be produced too dry. Vacuum filter cloth changes corrected this problem.

Annual core borings of the landfills have been performed to insure that the ongoing quality control results are accurate, and that what is actually placed is meeting all the minimum criteria. All results have shown that when the processed material is properly conditioned, placed and compacted, the permit guidelines are satisfied by landfill disposal using the Poz-O-Tec system. Typical data from these borings are shown in Table #I.

TABLE I
TYPICAL 28 DAY MATERIAL PROPERTIES

	Mill Creek	Cane Run
Unconfined Compressive Strength	75 psi	48 psi
Permeability	2.5×10^{-6}	1.01×10^{-6}

COAL WASTE UTILIZATION IN ARTIFICIAL REEF CONSTRUCTION

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ABSTRACT

The technology of coal combustion has improved greatly over the last few decades. One result of the elaborate emission control equipment required on coal plants to meet stringent air quality requirements is the large volumes of wastes that must be utilized or disposed of safely. Flue-gas desulfurization (FGD) sludge and fly ash may be produced at the rate of as high as 1,000 tons per day at a typical coal fired power plant. Waste disposal, especially in urban coastal areas, has become a major obstacle to conversion to coal combustion for generating electricity.

To assess a possible solution to the waste disposal problem, 500 tons of FGD sludge and fly ash were stabilized into blocks and placed in the ocean as an artificial reef. Previous laboratory investigations had indicated that no toxic chemical or physical effects should occur in the marine environment. After three years in the sea, the coal waste blocks support a diverse community of reef fish and invertebrates and have maintained their structural integrity. No adverse environmental effects have been detected.

Assuming that the coal waste blocks continue to be environmentally acceptable in the marine environment, the engineering and economic feasibility of this method of disposal should now be confirmed.

INTRODUCTION

The increase in the combustion of coal for generating electricity has resulted in the improvement in the technology required to burn coal cleanly. Electrostatic precipitators are able to remove as much as 99% of the particulate ash and flue gas desulfurization (FGD) scrubbers reduce the sulfur oxide emissions to acceptable levels. A problem, though, is created -- waste disposal. With a large coal-burning power plant producing more than 1000 tons of fly ash and FGD sludge each day, coastal metropolitan areas and areas with shallow ground water tables may lack sufficient land to safely dispose of these wastes. Ocean disposal could become an option if the materials can be stabilized to prevent any harmful effects to the marine environment.

In September, 1980, 500 tons of stabilized FGD sludge and fly ash, in the form of 15,000 hardened blocks (20 x 20 x 40 cm), were placed in the Atlantic Ocean as artificial reef substrate. Two years of prior laboratory and small-scale field investigations sought to detect any possible adverse effects to the marine community as well as any physical deterioration of the blocks themselves. With the manufacture of the 15,000 blocks and their placement in the sea, a three-year monitoring program continued to investigate the environmental acceptability of constructing artificial reefs with coal wastes.

Several types of biological surveys were conducted to compare the success of the coal waste reef with that of an existing reef of rock and building rubble. The fish community was sampled with traps to estimate population densities. Test organisms were placed at the reef to detect possible uptake of block components. The structural integrity of the blocks was monitored using ultrasonic and compressive testing. To understand any chemical changes occurring and to allow estimation of the lifetime of the blocks in the sea, the chemical composition of the blocks was determined throughout the study.

As these investigations continue to provide positive results, the engineering and, ultimately, the economic feasibility of disposing of coal wastes in the ocean as artificial reefs is now being determined.

BLOCK PRODUCTION AND REEF CONSTRUCTION

The 500 tons of FGD sludge and fly ash were obtained from two power plants, the Conesville station of Columbus and Southern Ohio Electric (CSOE) company and the Petersburg station of the Indiana Power and Light Company (IPALCO). The Conesville material was produced with a ratio of fly ash to FGD sludge of approximately 3:1 while the IPALCO material had a fly ash:sludge ratio of about 1.5:1. The chemical compositions of the components of each material as well as the stabilized blocks were determined and reported in Parker _et_ _al_., (1981).

The actual block production utilized standard concrete block industry equipment at a commercial plant in Pennsylvania (Figure 1). Very few modifications had to be made to accommodate the mixture of fly ash, sludge, lime and cement which was somewhat wetter then conventional concrete materials. The 15,000 blocks were all cured for 24 hours in a 'wet steam' kiln at about 65°C and finally trucked to a dock in New Jersey. A bottom opening barge transported the blocks to the project site (Figure 2) and released them to settle on the bottom in 20 m depth of water.

BIOLOGICAL INVESTIGATIONS

To determine the success of the coal waste blocks as an artificial reef substrate, it was necessary to measure colonization of the block surfaces by

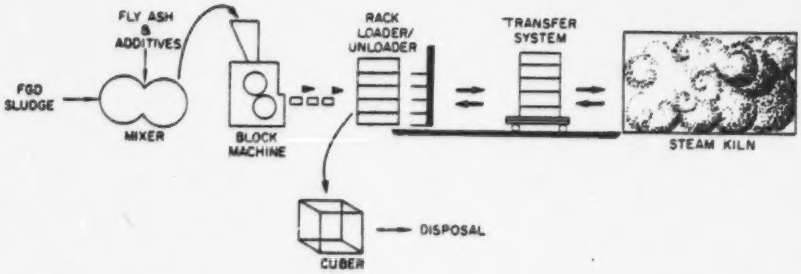


Figure 1. Schematic of block factory.

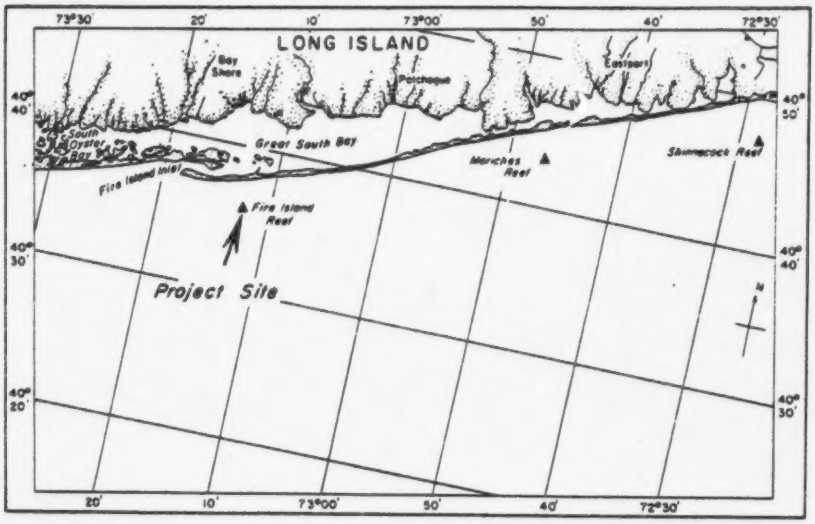


Figure 2. C-WARP project site.

epibenthic fauna and to determine the population of fish taking up residence in the "nooks and crannies" provided by the reef. Most importantly, analyses were made on reef organisms to detect possible uptake of toxic components from the blocks into the food chain. Special test bricks of coal waste were set out and returned to the laboratory periodically for identification of species and measurement of coverage (Figure 3 and Table 1). Close-up underwater photographs were also taken throughout the entire study period of the same areas on tagged blocks to document changes in the epibenthic community.

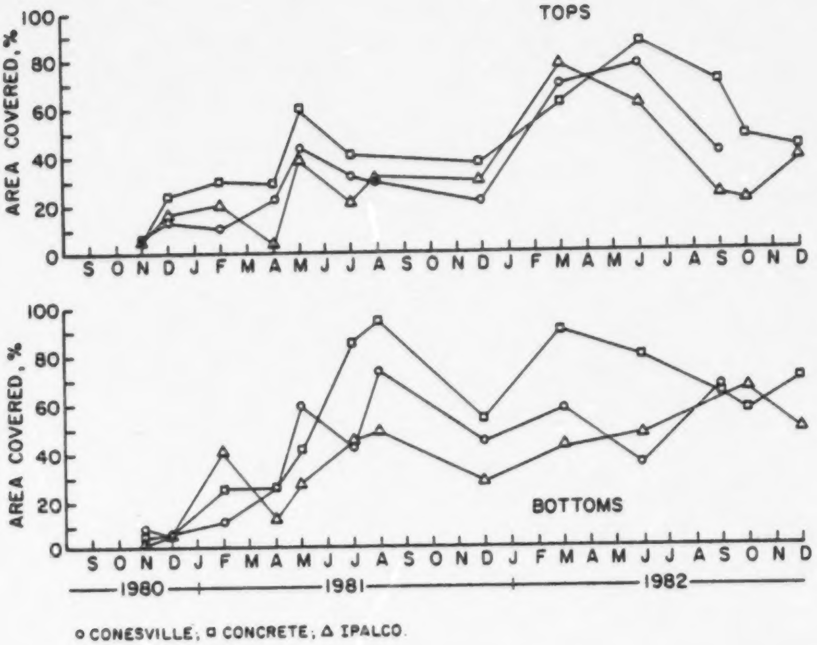


Figure 3. Coverage by epifaunal colonizers.

TABLE 1. EPIBENTHIC SPECIES GROWING ON SURFACES ON REEF BLOCKS, 1980-81.

Porifera
- Sponge, unident.

Colenterata
- *Metridium senile*
- *Clytia* sp.
- *Obelia dichotoma*
- *Sertularia cupressina*
- *Astrangia danae*

Bryozoa
- *Aetea* sp.
- *Bugula turrita***
- *Callopora aurita*
- *Cribulina punctata*
- *Electra hastingsae*
- *Electra pilosa*
- *Schizoporella unicornis***
- *Microporella* sp.

Mollusca
- *Anomia aculeata*
- *Anomia simplex*
- *Mytilus edulis***
- *Zirphaea crispata*
- *Spisula* sp.
- *Onchidoris* sp.
- *Acanthodoris pilosa*
- *Eubranchus exiguus*

Annelida
- *Harmothoe extenuata*,
- *Phyllodoce arenae*
- *Polydora socialis***
- *Syllis* sp.
- *Nereis grayi*
- *Sabellaria vulgaris***
- *Asabellides oculata*

Arthropoda
- *Balanus crenatus***
- *Caprella linearis*
- *Pokogeneia inermis*
- *Edotea triloba*
- *Unciola irrorata*
- *Cancer irroratus*
- harpacticoid copepods

Echinodermata
- *Asterias rubens*

Tunicata
- tunicate unident.

Fish traps were used to capture, tag, and release fish from both the coal waste reef and a long established fishing reef of rock and building rubble (Fire Island Reef) nearby. Because the area of the coal waste reef was similar to that of the part of Fire Island Reef being studied, a comparison of fish populations was conducted. The density of cunner on the coal waste reef increased from approximately 2.6 fish per m^2 in 1981 when the reef had been established for one year, to 8.5 fish per m^2 in 1982. During 1981 and 1982 the cunner population densities at the Fire Island Reef remained steady at approximately 7.5 fish per m^2. These results indicate that, in less than 2 years the coal waste reef had achieved "carrying capacities" for fish comparable to that of the existing artificial fishing reef.

Initial toxicity experiments in the laboratory tested sensitive marine organisms, a diatom and winter flounder eggs and larvae (Parker *et al.*, 1981), for adverse effects due to exposure to various concentrations of an elutriate made from powdered coal waste block material. The lack of significant effects was further investigated by field tests at the reef. Mussels, *Mytilus edulis*, widely used indicator organisms for toxic uptake studies, were placed at both the coal waste reef and at a separate control site. Again, analysis of mussel tissue showed no significant increases in potentially toxic trace elements after 9 months exposure on the reef. A laboratory experiment using mussels exposed to elevated concentrations of suspended coal waste material yielded similar results (Table 2). A spurious increase in copper was not seen at higher concentrations of coal waste. A doubling of iron concentrations in the mussels would not be expected to create any toxic effects. Fish were tagged during the three-year study period and recapture data indicate that these fish remained resident on the same reef throughout the complete study. Fish samples from the reef have been taken for tissue analysis of trace elements and should represent the most realistic case to detect possible uptake of block components.

TABLE 2. TRACE METAL CONTENT OF SOFT TISSUES FROM MUSSELS EXPOSED TO SUSPENSIONS OF POWDERED CONESVILLE COAL WASTE

CONCENTRATION OF COAL WASTE (mg/ℓ)	TRACE ELEMENTS (ppm, dry weight)						
	Cd	Cu	Fe	Mn	Ni	Pb	Zn
0 (control)	1.28	8.47	56.4	6.48	1.46	2.31	107.1
89	1.42	8.98	102.4	4.43	2.29	2.58	101.2
178	1.42	13.45	101.5	4.03	1.70	2.93	116.5
267	1.24	7.87	106.4	5.60	5.32	2.76	97.4

Values are averages of values for 5 replicates.
Underlined values are significantly different than controls at $p \leq 0.5$ level.

PHYSICAL AND CHEMICAL STABILITY

Although the blocks' surfaces successfully support a wide diversity of epibenthic colonizers, it is important to determine how quickly the blocks might deteriorate in the sea. Previous laboratory experiments indicated a gradual increase in compressive strength and density over time. After the placement of the reef, blocks were tagged for repeated ultrasonic (non-destructive) testing in the sea. Results have provided a good correlation between the series of ultrasonic data and compressive strength measurements when the blocks were finally returned to the laboratory after extended exposure at sea. Both sets of results indicate a gradual increase in strength seen in Figure 4 (Parker et al., 1982).

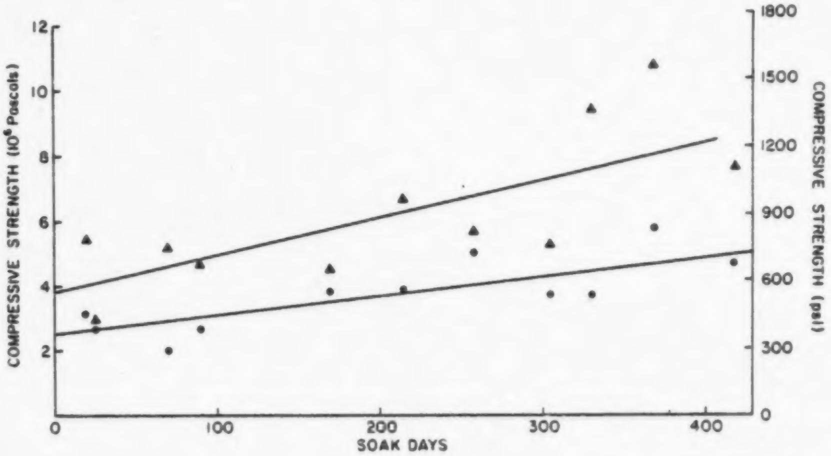
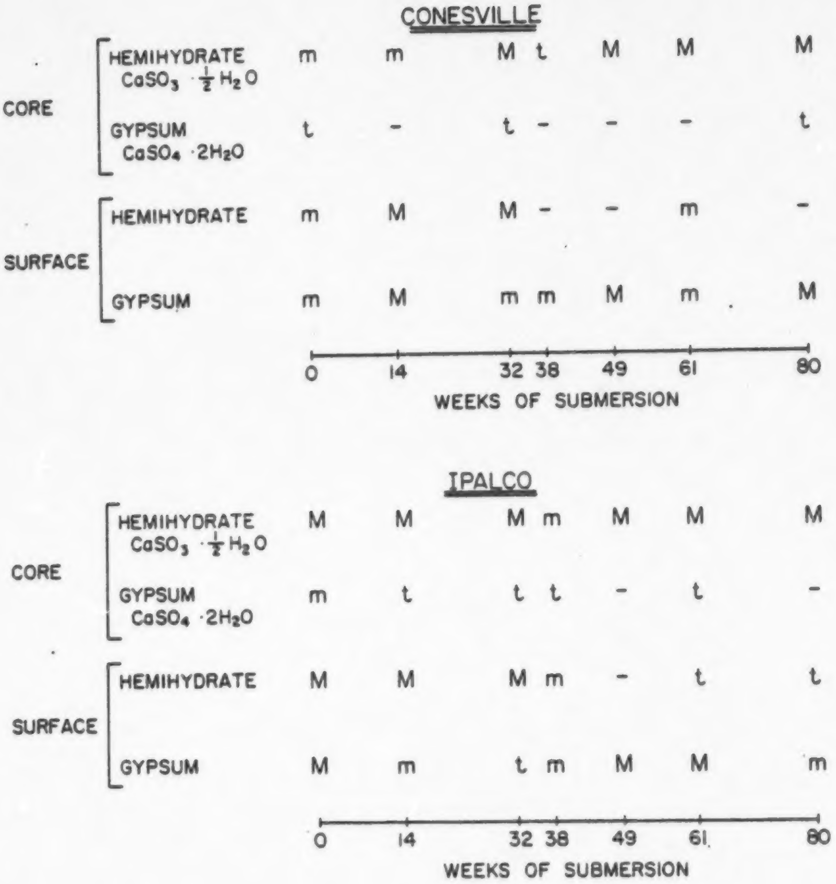


Figure 4. Compressive strength of Conesville (Δ) and IPALCO (O) blocks in the ocean.

The coal waste blocks undergo cementitious processes similar to those occurring in concrete. These reactions normally continue long after initial formation and it is important to understand what changes might occur due to protracted exposure to seawater at 3 atmospheres of pressure. Blocks returned from the sea were analyzed for any changes in mineralogy. As exposure time in the sea increased, mineralogical changes were found to occur in the surface layer of the blocks. In Figure 5, mineralogical results indicate that calcium sulfite hemihydrate and gypsum are major and minor components, respectively, in the core of the blocks. In the surface (1-2 cm) layer, however, a conversion has occurred making gypsum a major component. Although this reaction results in an overall positive volume change, the inherent porosity of the material accommodates any internal pressures that would otherwise develop.

Little, if any, change has occurred in the composition of minor and trace elements. However, significant surface losses of calcium by dissolution have been accompanied by an enrichment in magnesium. Laboratory



CONESVILLE

		0	14	32	38	49	61	80
CORE	HEMIHYDRATE $CaSO_3 \cdot \frac{1}{2}H_2O$	m	m	M	t	M	M	M
	GYPSUM $CaSO_4 \cdot 2H_2O$	t	-	t	-	-	-	t
SURFACE	HEMIHYDRATE	m	M	M	-	-	m	-
	GYPSUM	m	M	m	m	M	m	M

WEEKS OF SUBMERSION

IPALCO

		0	14	32	38	49	61	80
CORE	HEMIHYDRATE $CaSO_3 \cdot \frac{1}{2}H_2O$	M	M	M	m	M	M	M
	GYPSUM $CaSO_4 \cdot 2H_2O$	m	t	t	t	-	t	-
SURFACE	HEMIHYDRATE	M	M	M	m	-	t	t
	GYPSUM	M	m	t	m	M	M	m

WEEKS OF SUBMERSION

Figure 5. Mineralogical changes occurring in the blocks.

experiments have been conducted to measure the dissolution rate of calcium and, using these results in a model, to estimate the overall lifetime of the blocks. Preliminary results indicate that, after 30 years, the diffusion zone would only penetrate 2-3 cm into the blocks.

ENGINEERING ASPECTS

After preliminary tests making blocks with machines at research facilities of the Besser Company, Alpena, Michigan, the 500 tons of coal wastes were processed at a commercial concrete block plant utilizing standard "off-the-shelf" equipment. To move a step closer to the full-time disposal of coal wastes from a typical power plant using stabilized waste block production techniques for reef construction, the capability to handle as much as 1000 tons of fly ash and FGD sludge per day must be demonstrated. Figure 6 illustrates a simplified system for the ocean disposal of coal waste blocks. The process flow is indicated in Figure 7. Calculations were made for a hypothetical power station equipped with two 500 MW units, and a 45% load factor over a 30 year life. The details were taken from the EPRI FGD Sludge Disposal Manual, 2nd Ed. (1980). The fuel was an eastern coal with 3.5%

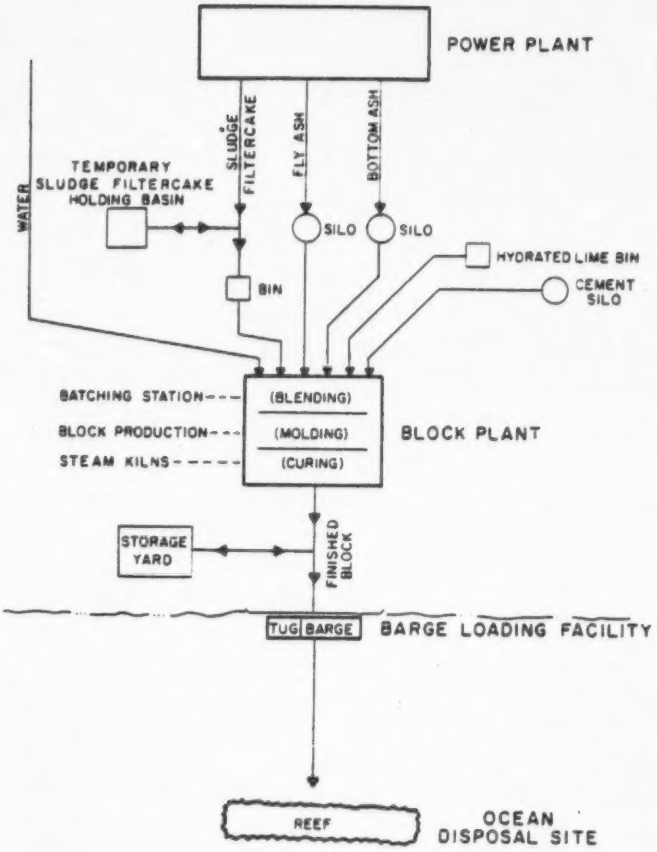


Figure 6. Block production design for ocean disposal.

sulfur and 12% ash content. Table 3 presents the estimations of the weights of coal waste requiring disposal daily.

For estimation of the capacity of machines to make blocks from sludge and fly ash, it has been assumed that 4% lime and 3% cement would be added to stabilize the waste mix and that the material would be consolidated to at least 1.76 g/cm³ (110 lb/ft³) as delivered from a block machine. The block size was that of a standard construction block, nominally 8 x 8 x 16 in (20 x 20 x 41 cm), but with an actual volume of 0.017 m³ (0.59 ft³) and weighing 29.4 kg (65.2 lb). Larger block molds are also readily available for use in conventional block machines.

The calculations of materials processing capacity (Table 4) indicate that a single V6-12, or two V3-12 block machines, operating at a rate of eight block-forming cycles per minute and with an efficiency of 90%, have the potential to fabricate the entire daily waste production of the power plant into blocks for placement in the ocean.

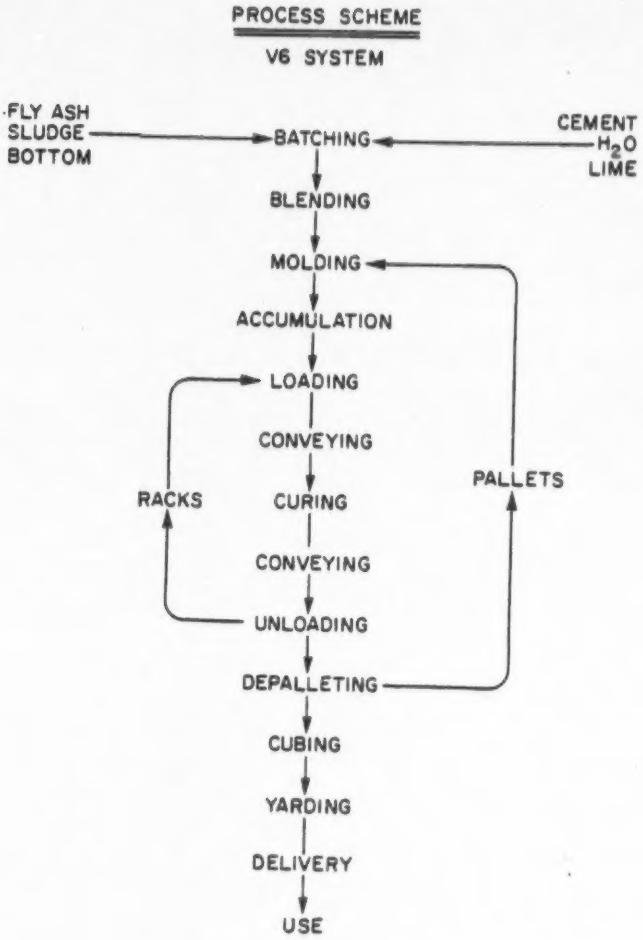


Figure 7. Process flow for coal waste block production.

Although the location and layout of an existing power plant would control how the block facility was constructed, a design that minimizes handling of the blocks in transfer to the ocean-going barge would greatly reduce costs.

REEF PLACEMENT ASPECTS

For a power plant producing 1000 tons of coal wastes per day, as many as 31,000 blocks could require placement in the ocean. Suitable sites must be chosen for the placement of the blocks if the benefits of the artificial reefs are to be fully realized. Economics factors will dictate the maximum distance from the loading facility but bottom type and accessibility to recreational fishermen will also be considered when regulatory agencies are asked to approve the designation of disposal/construction sites. The hard, sandy bottom and number of inlets off Long Island, New York make this area very suitable for artificial reef construction. With several hundred square miles of ocean bottom between the depths of 40 and 90 feet, sites can be chosen which would minimize interference with commercial fishing operations,

Table 3. ESTIMATION OF AVERAGE WASTE QUANTITIES FOR 3.5% SULFUR EASTERN COAL (FROM EPRI REPORT CS-2009 VOL. 2, NOV. 1981).

ASSUMPTIONS (for a two 500 MW unit generating station)

COAL

Btu/lb	12,500
Ash content, %	12
Fly ash/bottom ash	80/20
Sulfur content, %	3.5
Annual consumption @ 45% capacity factor, metric ton	1,430,000

AIR QUALITY CONTROL SYSTEM

Upstream fly ash removal efficiency, %	99
Overall fly ash removal efficiency, %	99.8
SO_2 scrubbing reagent	Lime
SO_2 removal, %	90
Stoichiometry (CaO/SO_2 removed), %	110
$CaSO_4 \cdot 2H_2O/CaSO_3 \cdot \frac{1}{2}H_2O$	50/50
Thickener underflow solids, %	35
Dewatered sludge solids, %	60

ASH PRODUCTION DAILY	METRIC TONS
Total ash produced	470
Total fly ash produced	376
Fly ash collected in precipitator	373

SLUDGE PRODUCTION DAILY

Fly ash collected in scrubber	3
$CaSO_4 \cdot 2H_2O$	132
$CaSO_3 \cdot \frac{1}{2}H_2O$	398
Weight of excess and unreacted lime reagent	108
Weight of dry sludge solids	642
Weight of water, at 60% solids content	427
Total weight of wet sludge, at 60% solids	1069

STABILIZATION ADDITIVES

Lime, 4%	41
Cement, 3%	30

WASTES TO BE PROCESSED DAILY	METRIC TONS
Fly ash + sludge solids + lime + cement	373 + 642 + 41 + 30
Total dry weight	1086
Water in sludge at 60% solids	427
Total wet weight	1513

TABLE 4. CALCULATION OF COAL WASTE BLOCK PRODUCTION RATES BY ONE V6-12 SUPERPAC OR TWO V3-12 VIBRAPAC MACHINES
(FOLLOWING ASSUMPTIONS IN TABLE 1)

Block unit size, nominal:	20.3 x 20.3 x 40.6 cm (8 x 8 x 16 in)	
Block unit volume:	0.017 m^3 (0.59 ft^3)	
Block forming cycle:	One 6-12 block machine	= 1 pallet of 6 units
		= 0.100 m^3 (3.56 ft^3)
	or two V3-12 block machines	= 2 pallets of 3 units
		= 0.100 m^3 (3.56 ft^3)
Block wet density:	1.76 g/cm^3 (110 lb/ft^3)	
Weight of 6 blocks:	176.7 kg (391 lb)	
Block forming rate:	8 cycles per minute	
Hourly block production:	176.6 x 8 x 60	= 84,805 kg/hr
		= 84.8 metric/hr
		(935 short ton/hr)
	At 90% efficiency	= 76.3 metric ton/hr
		(84.2 short ton/hr)
	Ash & sludge (dwb) process rate	= 76.3 x 71.7% solids
		= 54.7 metric ton/hr
		(60.3 short ton/hr)

From Table 1,

Coal wastes to be processed daily (dwb) = 1086 metric ton (1197 short ton)

Block production time per day $= \frac{\text{1086 metric ton/day}}{\text{54.7 metric ton/hr}} =$ 19.8 hours

prevent damage to shellfish areas, and would not "fill in the ocean." In fact, assuming the production of 31,000 blocks each day from a power plant and allowing a doubling of volume when released on the bottom due to random stacking, over 30 years of blocks could be contained in a reef covering 1 square mile and rising only 10 feet above the bottom. Water depth would control how high the reef could be constructed, due to potential hazard to navigation. Other potential uses of the blocks include rehabilitation of old sand mining areas and enhancement of fishing in areas that, historically, have had soft bottoms.

THE ECONOMIC VIABILITY OF COAL WASTE REEFS

This program has demonstrated the technical feasibility of processing power plant coal combustion by-products into blocks by using conventional concrete block making equipment. The environmental acceptability has been demonstrated in the laboratory and field investigations. To answer the big question on whether this concept is commercially viable, the Electric Power Research Institute commissioned Michael Baker, Jr., Inc. to do an evaluation of this technology from a hypothetical coal plant located in the New York Metropolitan area. The reef construction system would consist of four parts: (1) The block product plant with storage facilities; (2) a barge loading facility; (3) transportation to the reef site for placement, and (4) monitoring the site. The engineering-economic assessment indicated that a hypothetical 600-MW power plant would produce nearly 300,000 tons of by-products each year. The reef construction system would cost about $45/ton to build and operate. This cost is double the amount that most power plants spend on by-product disposal when land is available nearby. In highly urbanized areas where land is not available, the disposal costs are nearly the same. Therefore the potential for reef construction as a viable alternative to traditional waste disposal technologies looks promising and can be considered by coastal utilities in waste management planning.

The comparative economics of construction of coal waste artificial fishing reefs versus conventional land disposal practices would be drastically changed if a bill now in the 98th Congress becomes law. The proposed legislation (H.R. 3474) is called the National Fishing Enhancement Act of 1983. The legislation would establish national standards for the construction of siting of artificial reefs in the waters of the United States in order to enhance fishery resources and fishing opportunities. To stimulate the construction of artificial fishing reefs, the proposed law provides a tax credit equal to the excess costs of reef construction over the usual disposal costs. This bill if enacted would put artificial reefs on an equal footing with conventional disposal costs anywhere. The reason for the congressional interest in artificial reefs is that fishery products provide an important source of protein and industrial products for U.S. consumption, yet U.S. production annually falls $3.2 Billion short of satisfying demand.

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SOLID WASTE ENVIRONMENTAL STUDIES AT ELECTRIC POWER RESEARCH INSTITUTE

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ABSTRACT

Solid Waste Environmental Studies (SWES) at Electric Power Research Institute (EPRI) is a comprehensive research project aimed at generating predictive methods and the essential data bases to evaluate the effect of disposal and reuse of solid wastes produced from fossil fuel combustion and flue gas cleanup operations on groundwater quality. EPRI has developed detailed research plans and has initiated research in leaching chemistry, chemical attenuation mechanisms, groundwater transport processes, and the evaluation of existing geohydrochemical models. For the next three or four years, fundamental research in geochemistry and geohydrology is expected to yield quantitative data on release rates, transformation characteristics, and subsurface transport of inorganic solutes leached from waste. Results of the research will be integrated by improving or developing new predictive methods and by validating the results with data from operating facilities.

INTRODUCTION

Do constituents released from land disposal and from reuse of solid residues by electric utilities influence groundwater quality in the surrounding area? What factors control leachability of chemicals? What solid waste types are of concern? What is the environmental fate of leached solutes? When and to what extent should technologies be applied to contain leachates? EPRI's SWES project was initiated in 1982 to develop useful answers to these and other questions.

SWES is a comprehensive research project dealing with fundamental studies in geochemistry and subsurface hydrology to produce methods and associated data sets for use in evaluating the effects of disposal and the reuse of solid residues on groundwater quality. The solid residues most emphasized in the SWES research are coal fly ash, coal bottom ash, mixtures of the two, flue gas desulfurization (FGD) sludges with fly and bottom ash, and oil ash.

GOALS AND OBJECTIVES

The long-term or ultimate goal of the SWES project is to develop and validate methods (namely geohydrochemical models) for predicting the fate of inorganic solutes released to the subsurface environments from utility industry operations.

This long-term goal will be met by first conducting fundamental research to define the cause-and-effect relationships of greatest importance and of highest uncertainty to the development of predictive methods. Once the fundamentals in the research areas are well understood and data sets are properly developed, the results are to be integrated into predictive methods (models) and validated with field data.

The near-term concurrent research is to accomplish the following:

(a) Evaluate existing geohydrochemical models and computer codes and assemble an interim usable geohydrochemical model(s)

(b) Develop quantitative data on the leaching chemistry of solid residues

(c) Develop quantitative data on the chemical attenuation in the geologic environment of inorganic constituents

(d) Develop quantitative data on groundwater transport of solutes

(e) Develop mathematical descriptions of important leaching processes, chemical attenuation mechanisms, and hydrological transport factors in the migration of solutes in groundwaters.

RESEARCH PLANNING

EPRI's initial research into solid waste environmental studies began in 1977 when projects were initiated on (1) chemical characterization of coal fly ash (RP1061), (2) evaluation of reproducibility of EPA's extraction procedure as applied to utility solid wastes (RP1487), (3) comparison of the physical and chemical properties of solid wastes from coal combustion and coal gasification processes (RP1486), and (4) providing a statistical evaluation of the variability of a representative coal ash sample from an individual power plant (RP1620). Until late 1980, the degree of emphasis that should be placed on solid waste disposal and groundwater pollution potential was undetermined.

In 1981 a planning study was conducted to define research for the SWES project. This work was carried out through four regional meetings, simultaneous use of questionnaires addressed to about 260 individuals followed by a research planning workshop, and continuing discussions in SWES Advisory Committee meetings. As a result, research plans for individual SWES contracts have been developed to accomplish its objectives. The bases for this research effort are described in detail in two EPRI reports (Murarka 1982 and SAI 1982). Figure 1 provides a compartmentalized sketch defining the overall problem of estimating the release, transport, transformation, and fate of leached constituents. To date EPRI has initiated research contracts in the high-priority research areas of waste leaching chemistry, chemical attenuation of solutes in the geological environment, transport of solutes to groundwaters, evaluation of existing geohydrochemical models, and evaluation of groundwater sampling methods and related field measurements. Present plans call for an expenditure of about $21 million dollars over the next six years.

RESEARCH METHODOLOGY AND MANAGEMENT

SWES is divided into the following six parallel research efforts:

1) Identifying and quantifying the solutes present in the utility industry solid residues

2) Developing data on the leaching chemistry of these solutes from the solids

3) Developing data on the chemical attenuation of the solutes in the geological environments

4) Identifying and quantifying groundwater transport parameters for solute migration

5) Evaluating, improving, and validating methods (primarily geohydrochemical model[s]) that can predict the fate of solutes in land, surface, and groundwaters

6) Evaluating and testing groundwater sampling and related measurement methods for use in the field environments.

The implementation of research in the six categories employs a general strategy of evaluating the existing state of knowledge through critical literature reviews, a detailed specification of experimental designs, and laboratory and field experiments to develop the quantitative results. An integrated analysis of the data would then provide mathematical descriptions of cause-and-effect relationships for use in predictive methods. This analysis would also provide data values for input parameters to the predictive methods. While the fundamental research continues toward achievement of the long-term goal, an interim assemblage of geohydrochemical model(s) will provide usable predictive methods for the near-term.

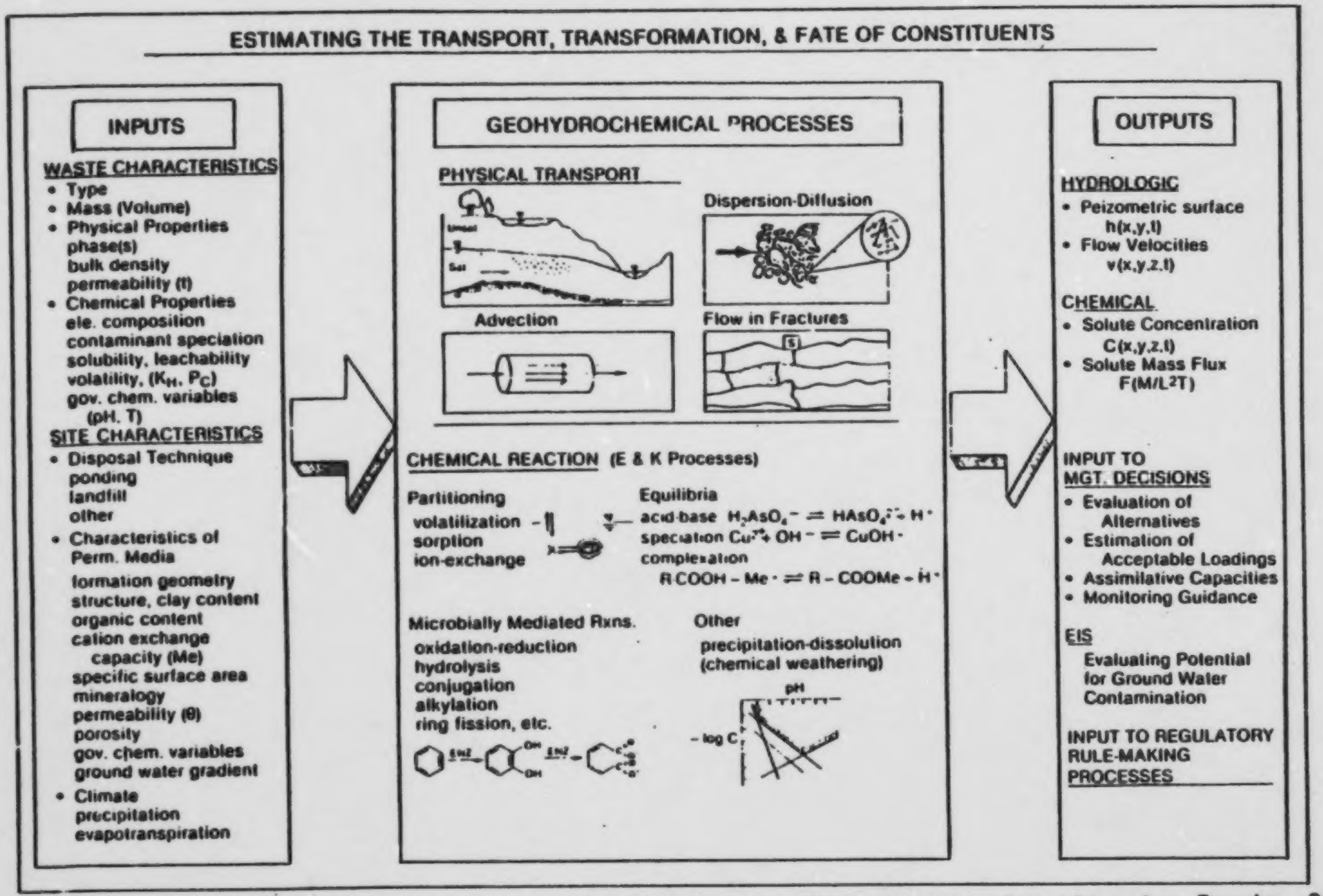


Figure 1. Estimating the Transport, Transformation, and Fate of Waste Constituents: Inputs, Outputs, and Example Processes*

*This figure is taken from EPRI Report EA-2415, Figure 1-2.

The management and technical direction of SWES contracts are accomplished primarily through the EPRI project manager. However, because of the nature and magnitude of this project, a technical management contractor is added to assist the EPRI project manager. In addition, an SWES Advisory Committee has been formed to advise the EPRI project manager on the SWES research. This advisory committee is composed of seven utility industry scientists, six scientists from academia and government agencies, two scientists from other EPRI divisions, and the EPRI project manager as the chairman.

RESEARCH CONTRACTS

SWES now has seven contracts to conduct the parallel research (fig. 2). A short description of the scope of work for these contracts is provided in the following sections.

Geohydrochemical Models Evaluation and Development

This was the first SWES contract awarded. The research in this effort has been divided into three phases:

- Phase I: The contractor is evaluating several models (computer codes) on geochemistry, geohydrology, and microbiological processes for their ability to predict the fate of constituents in groundwater. This phase of research is to be completed by March 1984.
- Phase II: The contractor will assemble interim usable geohydrochemical model(s) by collating the best methods identified during the Phase I research. The interim models connecting geochemical and geohydrological processes for modeling the fate of selected inorganic constituents will be tailored for near-term use by the industry. The Phase II research is to be completed by the end of 1986.
- Phase III: The results from research in chemical attenuation, leaching chemistry, and groundwater transport processes will be integrated in the improvement and field validation of geohydrochemical models. Present plans call for Phase III research to be completed by 1990.

Attenuation Rates, Coefficients, and Constants in Leachate Migration

Chemical interaction of solutes in solution with the surrounding soils and geological materials has profound effect on the migration of the leached constituents to groundwater. Therefore, the study of chemical attenuation has

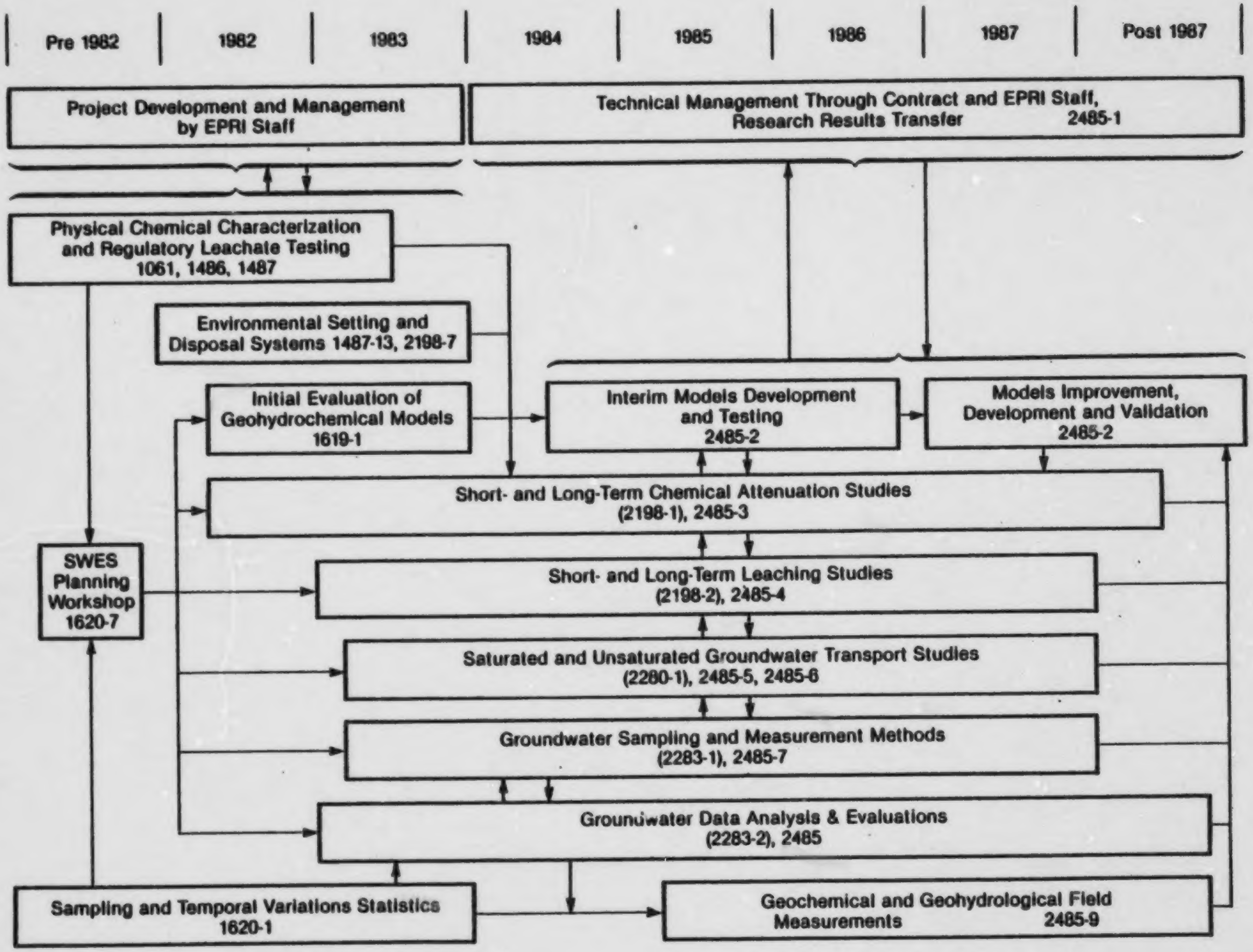

Pre 1982
1982
1983
1984
1985
1986
1987
Post 1987
Project Development and Management by EPRI Staff
Technical Management Through Contract and EPRI Staff, Research Results Transfer 2485-1
Physical Chemical Characterization and Regulatory Leachate Testing 1061, 1486, 1487
Environmental Setting and Disposal Systems 1487-13, 2198-7
Initial Evaluation of Geohydrochemical Models 1619-1
Interim Models Development and Testing 2485-2
Models Improvement, Development and Validation 2485-2
SWES Planning Workshop 1620-7
Short- and Long-Term Chemical Attenuation Studies (2198-1), 2485-3
Short- and Long-Term Leaching Studies (2198-2), 2485-4
Saturated and Unsaturated Groundwater Transport Studies (2280-1), 2485-5, 2485-6
Groundwater Sampling and Measurement Methods (2283-1), 2485-7
Groundwater Data Analysis & Evaluations (2283-2), 2485
Sampling and Temporal Variations Statistics 1620-1
Geochemical and Geohydrological Field Measurements 2485-9

Figure 2. Solid Waste Environmental Studies (SWES) Project Contracts & Schedule Environmental Physics & Chemistry Program, EPRI, September 1983.

been considered as the area that can produce the largest incremental improvements in methods for predicting leachate migration to groundwaters. Thus, this research is aimed at fundamental processes at the onset.

The research has been divided in two successive phases. In Phase I the researchers have compiled and collated the quantitative data from existing literature on chemical attenuation rates, constants, and coefficients for 21 inorganic elements (Table 1).

The critical evaluation of the literature in Phase I has included data derived from laboratory and field research. Emphasis has been placed on the acquisition of existing data on adsorption-desorption, precipitation-dissolution, and thermodynamic data on stability constants for the 21 inorganic elements. Thermodynamic data have been used to evaluate the coefficients and constants for chemical attenuation mechanisms applicable to leachates from utility industry solid residues. A critical review of the experimental or observational procedure used behind the literature data has been done prior to accepting the data values for this research. This review of literature and assembly of data was completed in July 1983.

TABLE 1. INORGANIC ELEMENTS FOR CHEMICAL ATTENUATION STUDIES

Aluminum
Antimony
Arsenic
Barium
Beryllium
Boron
Cadmium
Chromium
Copper
Fluoride
Iron
Lead
Manganese
Mercury
Molybdenum
Nickel
Selenium
Sodium
Sulfate
Vanadium
Zinc

In Phase II the contractor designed and started laboratory experiments (on solubilities, solubility products, and kinetics of precipitation-dissolution of solid phases) to produce data on the solubility-limited concentrations of selected elements in the natural environment. Additional experiments will be conducted to measure intrinsic adsorption constants, quantify the effects of competing ions and complexing ligands on the adsorption-desorption characteristics, and to quantify adsorption-desorption on mineral mixtures and natural soils for single and multiple electrolytes. These experiments will employ batch, column, and controlled field set-ups to simulate the complexity of the environment. Attenuation rate constants will be established by summarizing the results. This contract research is expected to be complete by the end of 1987.

Solid Waste Leaching Studies

What factors (e.g., composition of waste and environmental conditions) control leachability? What constituents can be leached from wastes? Under what conditions? In which chemical form? How does one determine the rate of leaching of a constituent for field conditions? Answers to most of these questions lie in understanding the chemical mechanisms. Earlier research by Hulett et al. (1981) and by Turner et al. (1982) has identified chemical forms of elements and distribution of trace elements found on the different solid phases of fly ash samples.

This work has provided a firm basis for research on chemical mechanisms for waste leaching. Therefore, the objectives of the planned leaching studies are to collate available data, critically evaluate the methods used in producing the data, and develop quantitative results on the leaching chemistry of solid residues. The research on leaching studies is divided into two phases. In Phase I the researchers are compiling quantitative data from existing literature, conducting feasibility experiments, and designing experiments for Phase II. All batch, column, sequential, and other leaching methods used to produce quantitative data on the chemical composition of leachates are being critically examined to identify relevant variables that are particularly important in controlling the leaching behavior of the fossil fuel combustion wastes. These variables could include factors such as occluded liquors, pH and quality of natural waters that such wastes come in contact with, solid-liquid ratio, permeability of wastes, age of wastes, contact time between waste solids and liquids, and the solid phases present. This phase of research is to be completed by April 1984.

In Phase II the researchers will conduct laboratory and controlled field experiments as a means of developing quantitative data on the leaching rates of constituents from utility solid wastes. These experiments will be conducted to include a broad range of waste materials and environmental conditions such that the results can be useful to a large segment of the utility industry. Phase II research is expected to be complete in 1986.

Groundwater Transport Studies

Solutes generally migrate by groundwater flow. The groundwater flow and transport calculations are based on mathematical equations using Darcy's formulation during the past 50 years. But the calculations on solute transport by groundwater are relatively new, perhaps less than 10 years old. The solute transport problem has two major processes: a chemical interaction phenomenon and a physical dispersion phenomenon. The groundwater transport studies relate to the physical dispersion phenomenon. Two contracts have been let out to conduct field experiments on groundwater transport of solutes in the porous media.

The research is divided in two phases. In Phase I the researchers are compiling and evaluating the existing data on physical-hydrological transport factors used to predict the fate of solutes in groundwaters. Based on this critical review, tracer experiments will be designed for several field sites to measure the transport parameters in the development of plumes. The Phase I research is to be completed by June 1984. In Phase II experiments are to be conducted for at least three sites for a period of three or more years to estimate the hydrodynamic dispersion parameters as a function of environmental variables and the solutes of interest. The groundwater transport research is expected to be complete by the end of 1987.

Groundwater Sampling Methods and Related Field Measurements

Because the ultimate goal of SWES is to produce methods for predicting the fate of solutes in the environment, reliable input data (e.g., soil and waste permeability, groundwater flow velocity, and predictive groundwater flow directions) are needed. The research in this contract establishes the availability and performance of field methods for sampling geochemical and groundwater environments specifically.

This research is divided into two successive phases. In Phase I the contractor is to compile and critically evaluate existing geochemical and hydrologic field sampling procedures. This evaluation will focus on at least the following three categories of sampling methods:

1) Groundwater sample collection, handling procedures, and instruments

2) Soil and waste sample collection, handling procedures, and instruments specifically for quantifying hydrologic and geochemical properties

3) Geophysical methods for geologic and subsurface hydrogeologic measurements.

The contractor is to prepare a manual containing details on field sampling procedures to define the strengths, weaknesses, and applicability to the utility industry. The Phase I research is to be completed by August 1984.

In Phase II the contractor will develop and field test procedures for groundwater sampling problems identified in Phase I. These tests will be repeated several times to establish reproducibility, accuracy, and ease of use for utility industry applications. The Phase II research is expected to be complete in 1987.

Groundwater Data Evaluation

To properly direct the applied research in SWES, a contract has been issued to analyze data from a full-scale field monitoring project conducted for the U.S. EPA and individual utility companies. The information from these studies includes data on groundwater quality, hydrology, soils and related geological measurements, and waste disposal systems. The contractor will analyze these data to identify hypotheses that should be tested in the SWES research on chemical attenuation, leaching chemistry, and groundwater transport of solutes. This data analysis is to be completed by July 1984.

Geochemical, Geohydrological, and Groundwater Quality Measurements

This research is planned for the later part of SWES. Intensive measurements at a selected number of sites will be made to develop a complete data base for the testing and validation of predictive methods developed in the SWES project. This contract is expected to begin in 1985 at the earliest and may continue for approximately four years. Detailed sampling plans for this contract will be specified prior to the beginning of the measurements.

RESEARCH RESULTS

Since this research began in 1977 and the initial phase of the current SWES research started in 1982, reports are now available that offer usable results. In this section a summary of the technical results from several EPRI reports has been prepared. For complete details, readers are referred to the reports sited in this paper.

EXTRACTION PROCEDURE AND UTILITY INDUSTRY WASTES

Two EPRI reports (Rose et al. 1981 and Eynon et al. 1983) deal with the topic of reproducibility of results (concentration of eight elements: As, Ba, Cd, Cr, Pb, Ag, Se, and Hg) in the extracts from using the extraction procedure (EP) specified by the U.S. EPA as applied to fly ash, bottom ash, and scrubber sludge. The concentration of silver in the EP extract was found to be below detection limit most of the time; therefore, no statistical analysis was performed on the data for silver.

The solid samples used in this research were: (1) dry fly ash--alkaline (DFA--ALK), (2) dry fly ash--acidic (DFA--AC), (3) wet bottom ash--alkaline (WA--ALK), (4) wet bottom ash--acidic (WA--AC), and (5) scrubber sludge (SS). Four different laboratories were used to extract four samples for each of the five waste types using the EPA's extraction procedure of December 1978. Each extract liquid was split into eight aliquots by each laboratory. Each aliquot sample was analyzed for the eight elements using two different analytic procedures. Numerical results obtained from this research are summarized in Table 2. Based on this research the following conclusions can be drawn:

- Results on the concentration of the eight elements were found to be hundreds of percent apart between different laboratories.

TABLE 2: SUMMARY STATISTICS AND APPORTIONED VARIATION IN LEACHATE COMPOSITION*

Element	Analysis	Prep	Waste	N Samp	N Used	Geom Mean (ppb)	Minimum (ppb)	Maximum (ppb)	Percent of Total Variance Inter Lab Extr	Intra Lab Extr	Inter Lab Anal-I	Inter Lab Anal-II	Intra Lab Anal
Arsenic	FURNACE AA	EP	DFA-AC	128	128	7.14	0.63	46.00	32	2	51	9	6
			DFA-ALK	128	128	7.07	0.63	110.00	4	44	31	7	13
			SS	128	128	77.22	10.00	220.00	47	12	37	2	2
	FLAME AA	EP	DFA-AC	128	128	6.21	1.00	140.00	35	2	14	17	32
			DFA-ALK	128	128	4.58	0.17	147.23	10	33	4	36	17
			SS	128	128	42.98	6.55	288.74	29	1	10	23	38
Barium	FURNACE AA	EP	DFA-AC	128	128	69.96	1.43	650.00	0	0	82	5	12
			DFA-ALK	128	128	129.32	0.63	3400.00	0	3	73	18	6
			WA-AC	128	128	43.23	1.43	1300.00	0	7	65	12	15
			WA-ALK	128	128	181.79	0.49	1600.00	2	0	88	0	11
			SS	128	128	83.44	1.26	730.00	1	0	90	0	10
	FLAME AA	EP	WA-ALK	128	127	890.54	24.86	32000.00	6	0	63	16	15
Cadmium	FURNACE AA	EP	DFA-AC	128	128	95.81	23.00	240.00	28	1	45	7	20
			DFA-ALK	128	128	0.19	0.01	48.00	0	29	41	2	28
			WA-AC	128	128	0.37	0.03	11.00	12	12	56	2	17
			WA-ALK	128	128	1.56	0.28	22.00	7	6	65	11	10
			SS	128	128	7.09	0.58	62.31	13	5	70	4	8
	FLAME AA	EP	DFA-AC	126	128	126.33	38.03	240.00	29	0	32	13	26
Chromium	FURNACE AA	EP	DFA-AC	128	128	9.01	2.50	31.69	0	28	4	19	50
			DFA-ALK	128	128	536.74	331.57	830.00	11	39	13	13	24
			WA-AC	128	128	1.09	0.11	412.37	0	71	9	10	10
			WA-ALK	128	128	2.72	0.16	290.00	42	18	2	22	16
			SS	128	128	28.32	2.61	110.00	70	24	3	0	3
	FLAME AA	EP	DFA-ALK	128	128	590.10	195.38	1000.00	3	65	15	13	4
Mercury	FLAME AA	EP	SS	128	128	0.73	0.05	59.00	53	4	0	24	19
Lead	FURNACE AA	EP	DFA-AC	128	128	1.04	0.14	9.00	12	0	54	7	27
			SS	128	128	0.88	0.18	5.44	6	0	32	25	38
Selenium	FURNACE AA	EP	DFA-AC	128	128	24.11	4.63	70.00	23	2	34	6	36
			DFA-ALK	128	128	51.88	23.57	121.62	16	34	11	22	17
			SS	128	128	39.08	15.69	73.00	0	15	13	23	49
	FLAME AA	EP	DFA-AC	128	128	5.25	0.05	62.00	4	0	86	3	6
			DFA-ALK	128	128	20.37	0.30	89.00	6	15	58	14	7
			SS	128	128	18.60	0.60	68.00	2	0	72	4	21

* The data in this table are taken from EPRI Report EA-3181, Tables 3-2 and 3-5.

- Reproduciblity differed from element to element, waste type, and the analytic technique used.

- The interlaboratory variation in results was generally higher than the intralaboratory variation.

TRACE ELEMENTS AND THEIR CHEMICAL FORMS

Two EPRI reports (Hulett et al. 1981 and Turner et al. 1982) deal with the general subject of ash chemistry. The objective for Part I in this research was to establish the distribution of trace elements in the various solid phases present in fly ash. By using physical and chemical methods, researchers first separated the fly ash samples into three phases: (1) glass (approximate composition Al-Si-O), (2) mullite-quartz (approximate composition $3(Al_2O_3).2(SiO_2)$), and (3) magnetic spinel (approximate composition Fe_2AlO_4). The distribution of trace elements in each of the three phases was then defined. The research results show:

- Glass is typically the most abundant solid phase of the non-magnetic alluminosilicates.

- Trace elements distribution analyses show that a distinct propensity occurs in specific phases:

  -- The glass phase seems to contain almost all of the alkali, alkaline earth, and rare earth elements, as well as the majority of As, Pb and Se.

  -- Small amounts of trace elements are associated with the mullite-quartz phase with the exception of V, Cr, Ga, Zr, Fe, and Ti--all of which seemingly occupy the +3 or +4 valence sites in the crystalline structure.

  -- The magnetic-spinel phase seems to be largely enriched in the first row transition elements (V, Cr, Mn, Fe, Co, Ni, Cu, and Zn).

The objective of research in Part II was to identify patterns in the coal ash composition and its leaching behavior in actual and laboratory-simulated disposal experiments. Arsenic was selected for a more complete study on leachability and chemical speciation in leachates. The important results reported by the researchers (Turner et al. 1982) are summarized below:

- Ash sluicing-ponding systems that are neutral to alkaline in pH appear to show appreciable solubilization of elements (e.g., S, As, Se, Cr, and Mo) with the anionic aqueous speciation. In contrast, the sluicing-ponding systems with pH in the acidic range appear to result in high solubilization of most cationic elements (particularly Fe, Mn, Cu, Zn, and Cd).

- The study of As indicated that in controlled pH experiments the As (v) varied systematically with pH variations and the As (III) concentration remained nearly constant over a wide range of pH (3 to 12).
- A considerable amount of variablility among fly ashes with respect to quantity and to the oxidation state of arsenic released was observed in the laboratory leaching experiments.

COMPARISON OF SOLID WASTES FROM COAL COMBUSTION AND PILOT COAL GASIFICATION PLANTS

The research results from this work are given in an EPRI report prepared by Turner and Lowry (1983). The researchers principally focused on developing data on the physical and chemical characteristics of solid wastes from conventional coal-burning power plants and solid wastes from coal gasification in the CE-PDU plant. Bulk composition results of the waste samples analyzed are shown in Table 3 and the EP extract composition results are shown in Table 4. The important results from this contract research are summarized below:

- Slag from the gasifier and slag from the cyclone-fired combustion unit were morphologically similar and predominantly consisted of a dense shards of specific surface area of less than 1 m^2/g. In contrast, the bottom ash from a pulverized-coal combustion plant predominantly contained round vesicular cinders of specific surface area between 1 and 3 m^2/g.
- The two slags did not show any crystalline structures, but mullite crystals were found in the fly ash from the pulverized coal combustion power plant.
- Based on aqueous extraction and inorganic elements analyses according to the EPA extraction procedure, none of the combustion or the gasification solid wastes studied would be classified as hazardous on the basis of the toxicity criteria of the Resource Conservation and Recovery Act regulations.
- The gasification slag showed a more acidic reaction in continuous-flow column extractions and thus may leach some constituents at a higher rate than combustion slag and bottom ash.
- Although some differences exist between the wastes examined, the gasification slag appears to be nearly as inert chemically as slag from the cyclone furnace.

TABLE 3. ELEMENTAL COMPOSITION OF COMBUSTION AND GASIFICATION SOLID WASTES PRODUCED FROM PITTSBURGH COAL*

	Mitchell Station**		CE-PDU***	Merrimack Station****	
	Fly Ash	Bottom Ash	Slag	Slag	Fly Ash ******
			%		
Si	20.5	21.6	19.2	21.4	18.0
Al	11.6	11.9	14.9	12.6	11.2
Fe	10.2	14.9	14.4	12.2	13.1
Ca	3.09	2.38	1.32	5.52	3.72
Na	0.56	0.55	0.87	0.94	1.34
Mg	0.45	0.46	0.31	1.62	0.73
K	1.19	1.10	1.32	1.03	1.58
Ti	0.62	0.56	0.59	0.60	0.58
S	0.80	0.18	0.31	0.07	0.91
C	7.04	0.58	0.80	0.16	4.6
			μg/g		
Ag	<1	<1	<2	<2	<2
As	93.	3.2	3.6	≤1.8	393.
Ba	743.	551.	1470.	1253.	1400.
Cd	0.82	<0.1	0.18	0.64	4.2
Cr	113.	87.	4750.	126.	298.
Cu	68.	44.	68.	69.	431.
Hg	0.10	0.006	0.011	0.008	0.04
Mn	188.	176.	273.	205.	182.
Ni	129.	122.	228.	98.	158.
Pb	50.	8.	4.9	8.1	309.
Se	12.	N.D.******	1.0	2.2	N.D.
Zn	92.	32.	28.	32.	460.

* This table is taken from EPRI Report EA-2867 Table 3.4.

** Pulverized coal furnace.

*** Entrained-flow gasifier (low Btu), collected June 3, 1980.

**** Wet-bottom slagging furnace.

***** Recycled to furnace at this power plant.

****** Not detected.

TABLE 4. RESULTS OF EPA'S EXTRACTION PROCEDURE APPLIED TO SOLID WASTES PRODUCED FROM PITTSBURGH COAL* (concentrations are in µg/L except for $SO_4^=$ which is in mg/L)

	Mitchell Station**		CE-PDU****	Merrimack Station***		Drinking Water Criteria
	Fly Ash	Bottom Ash	Slag	Slag	Fly Ash******	
Ag	0.06	<0.01	<0.05	<0.05	0.12	50
As	9.2	0.88	0.55	0.09	63.	50
Ba	145.	85.	58.	2.5	100.	1000
Cd	12.	0.12	0.06	0.11	69.	10
Cr	87.	1.3	1.45	<0.1	1.5	50
Cu	106.	1.4	5.1	0.74	290.	1000
Fe	362.	144.	326.	275.	22.	300
Hg	0.014	0.011	0.058	0.004	0.04	2
Mn	1700.	244.	4.45	1.3	580.	50
Ni	188.	11.	180.	0.28	210.	--
Pb	4.35	< 0.06	0.53	0.14	3.1	50
Se	2.8	< 0.05	<1	<0.5	11.	10
Zn	160.	5.7	5.2	0.61	1500.	5000
$SO_4^=$	940.	16.2	6.24	--	--	250
Acid, mL/g	1.2	0.15	0	0.01	1.2	
Final pH	5.0	5.0	4.6	4.8	5.1	

* This table is taken from EPRI Report EA-2867 Table 3.8.

** Pulverized coal furnace.

*** Wet-bottom slagging furnace.

**** Entrained flow gasifier (Low Btu), collected June 3, 1980.

***** Recycled to furnace at this power plant.

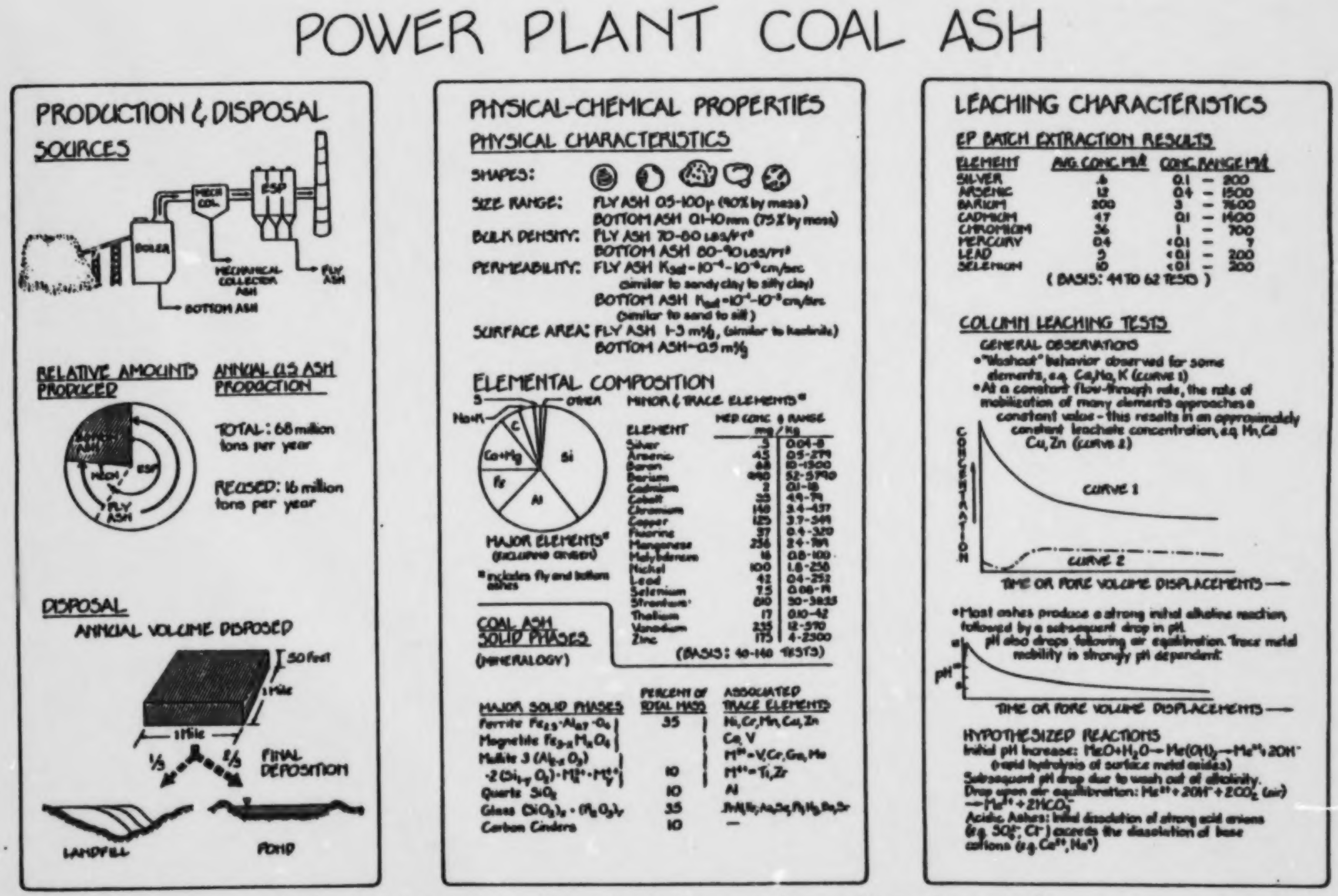
STUDY OF TIME-VARIABILITY OF ELEMENT CONCENTRATIONS IN POWER PLANT ASH

This research contract was initiated to study the uncertainty in ash composition due to representativeness of the ash samples subjected to the extraction test. This study is critical to answering: "What constitutes a representative waste sample, and how variable is the waste stream over time?" The research has been conducted in two parts. Results from the Part I pilot study are given in an EPRI report prepared by Switzer, Eynon, and Holcombe (1983). Part II research has also been completed and the report is in preparation. The Part I research consisted of obtaining samples of coal and coal ashes from a southwestern power plant for a 30-day period. The samples were chemically analyzed for 14 inorganic elements. The elemental concentration data have been subjected to statistical analyses to estimate the time variability in chemical composition of coal ashes. Results from this research are summarized below:

- Time, sampling, sample preparation, and sample analysis are the four components providing sizable contributions to the total variability in ash composition.
- The contribution of the four components to total variability differed in degree depending upon the chemical element and method of analysis.
- The correlation between plant process variables (feed coal composition, boiler temperature, boiler oxygen, and electrostatic precipitator air temperature) and the ash chemical composition showed very little association. This lack of correlation may be due to the relatively constant chemical composition of feed coal during this research.

PHYSICAL-CHEMICAL CHARACTERISTICS OF UTILITY SOLID WASTES

The primary purpose of this research was to compile existing data on waste physical properties, chemical composition, and leachate composition for estimating the physical-chemical composition of solid wates. The EPRI report by Summers, Rupp, and Gherini (1983) provides detail on the results. A large quantity of the data compiled are on coal ash and its leachates. These data are for 19 minor and trace elements. Relatively small amounts of data are on physical properties of ash and flue gas desulfurization (FGD) sludge, chemical composition of FGD sludge and leachates, and chemical composition of oil ash and geothermal wastes. The results of this research are summarized in Figures 3 and 4.



TOTAL: 68 million tons per year

REUSED: 16 million tons per year

SHAPES:

SIZE RANGE: FLY ASH 0.5-100μ (90% by mass)
BOTTOM ASH 0.1-10 mm (75% by mass)

BULK DENSITY: FLY ASH 70-80 LBS/FT³
BOTTOM ASH 80-90 LBS/FT³

PERMEABILITY: FLY ASH K_{sat} - 10^{-4} - 10^{-6} cm/sec (similar to sandy clay to silty clay)
BOTTOM ASH K_{sat} - 10^{-1} - 10^{-3} cm/sec (similar to sand to silt)

SURFACE AREA: FLY ASH 1-3 m²/g, (similar to kaolinite)
BOTTOM ASH - 0.5 m²/g

*includes fly and bottom ashes

MINOR & TRACE ELEMENTS*

ELEMENT	MED CONC. mg/kg	RANGE mg/kg
Silver	.5	0.04-8
Arsenic	45	0.5-279
Boron	68	10-1300
Barium	990	52-5790
Cadmium	2	0.1-18
Cobalt	35	4.9-79
Chromium	160	3.4-437
Copper	125	3.7-349
Fluorine	37	0.4-320
Manganese	236	2.4-789
Molybdenum	16	0.8-100
Nickel	100	1.8-258
Lead	42	0.4-252
Selenium	7.5	0.08-19
Strontium	810	30-3855
Thallium	17	0.10-42
Vanadium	235	12-570
Zinc	175	4-2300

(BASIS: 40-140 TESTS)

MAJOR SOLID PHASES	PERCENT OF TOTAL MASS	ASSOCIATED TRACE ELEMENTS
Ferrite $Fe_{2.3} \cdot Al_{0.7} \cdot O_4$	35	Ni, Cr, Mn, Cu, Zn
Magnetite $Fe_{3-x}M_xO_4$		Co, V
Mullite $3(Al_{2-x}O_3) \cdot 2(Si_{1-y}O_2) \cdot M_x^{3+} \cdot M_y^{4+}$	10	M^{3+} = V, Cr, Ga, Mo M^{4+} = Ti, Zr
Quartz SiO_2	10	Al
Glass $(SiO_2)_x \cdot (M_2O_3)_y$	35	Pb, Ni, Ag, As, Se, Hg, Ba, Sr
Carbon Cinders	10	—

EP BATCH EXTRACTION RESULTS

ELEMENT	AVG. CONC. μg/l	CONC. RANGE μg/l
SILVER	.6	0.1 — 200
ARSENIC	12	0.4 — 1500
BARIUM	200	3 — 7600
CADMIUM	47	0.1 — 1400
CHROMIUM	36	1 — 700
MERCURY	0.4	<0.1 — 7
LEAD	5	<0.1 — 200
SELENIUM	10	<0.1 — 200

(BASIS: 44 TO 62 TESTS)

COLUMN LEACHING TESTS

GENERAL OBSERVATIONS

- "Washout" behavior observed for some elements, e.g. Ca, Na, K (curve 1)
- At a constant flow-through rate, the rate of mobilization of many elements approaches a constant value - this results in an approximately constant leachate concentration, e.g. Mn, Cd, Cu, Zn (curve 2)

- Most ashes produce a strong initial alkaline reaction, followed by a subsequent drop in pH.
- pH also drops following air equilibration. Trace metal mobility is strongly pH dependent.

HYPOTHESIZED REACTIONS

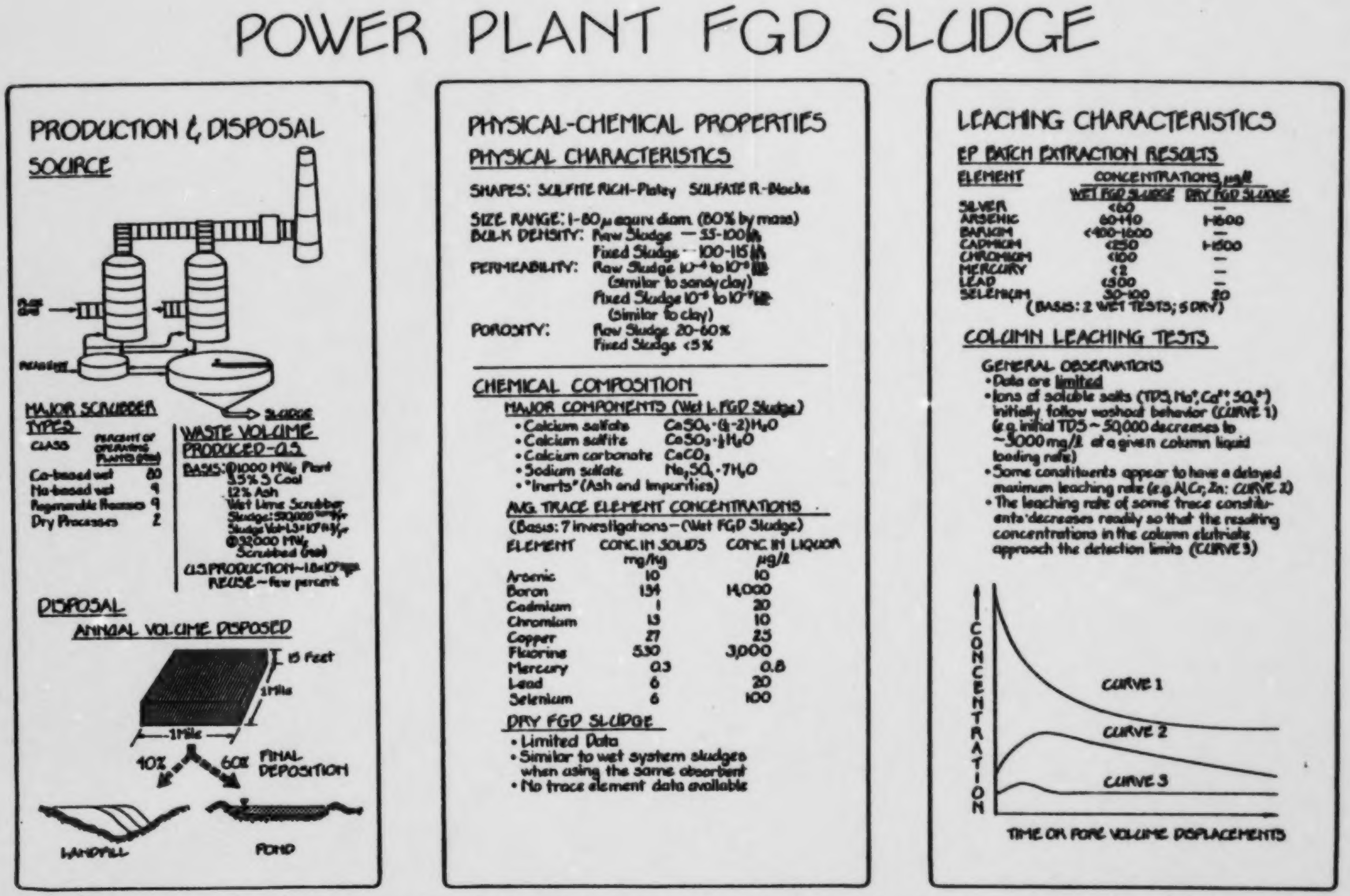
Initial pH increase: $MeO + H_2O \rightarrow Me(OH)_2 \rightarrow Me^{++} + 2OH^-$ (rapid hydrolysis of surface metal oxides)

Subsequent pH drop due to wash out of alkalinity.

Drop upon air equilibration: $Me^{++} + 2OH^- + 2CO_2\ (air) \rightarrow Me^{++} + 2HCO_3^-$

Acidic Ashes: Initial dissolution of strong acid anions (e.g. $SO_4^=$, Cl^-) exceeds the dissolution of base cations (e.g. Ca^{++}, Na^+)

Figure 3. Summary of Physical-Chemical Characteristics of Power Plant Coal Ash*
*This figure is taken from EPRI Report EA-3236, Figure 7-1.



POWER PLANT FGD SLUDGE

PRODUCTION & DISPOSAL

SOURCE

MAJOR SCRUBBER TYPES

CLASS	PERCENT OF OPERATING PLANTS (1980)
Ca-based wet	80
Na-based wet	9
Regenerable Processes	9
Dry Processes	2

WASTE VOLUME PRODUCED–U.S.

BASIS: ①1000 MWe Plant
3.5% S Coal
12% Ash
Wet Lime Scrubber
Sludge: 510,000 tons/yr
Sludge Vol–1.3×10⁷ ft³/yr
②32000 MWe Scrubbed (1980)

U.S. PRODUCTION–1.8×10⁷ tons/yr
REUSE–few percent

DISPOSAL

ANNUAL VOLUME DISPOSED

PHYSICAL-CHEMICAL PROPERTIES

PHYSICAL CHARACTERISTICS

SHAPES: SULFITE RICH–Platey SULFATE R.–Blocks

SIZE RANGE: 1-80μ equiv. diam. (80% by mass)
BULK DENSITY: Raw Sludge — 55-100 lb/ft³
Fixed Sludge — 100-115 lb/ft³
PERMEABILITY: Raw Sludge 10⁻⁴ to 10⁻⁵ cm/sec (similar to sandy clay)
Fixed Sludge 10⁻⁵ to 10⁻⁷ cm/sec (similar to clay)
POROSITY: Raw Sludge 20-60%
Fixed Sludge <5%

CHEMICAL COMPOSITION

MAJOR COMPONENTS (Wet L. FGD Sludge)

- Calcium sulfate $CaSO_4 \cdot (\frac{1}{2}\text{-}2)H_2O$
- Calcium sulfite $CaSO_3 \cdot \frac{1}{2}H_2O$
- Calcium carbonate $CaCO_3$
- Sodium sulfate $Na_2SO_4 \cdot 7H_2O$
- "Inerts" (Ash and Impurities)

AVG. TRACE ELEMENT CONCENTRATIONS

(Basis: 7 investigations–(Wet FGD Sludge)

ELEMENT	CONC. IN SOLIDS mg/kg	CONC. IN LIQUOR μg/l
Arsenic	10	10
Boron	134	14,000
Cadmium	1	20
Chromium	13	10
Copper	27	25
Fluorine	530	3,000
Mercury	0.3	0.8
Lead	6	20
Selenium	6	100

DRY FGD SLUDGE

- Limited Data
- Similar to wet system sludges when using the same absorbent
- No trace element data available

LEACHING CHARACTERISTICS

EP BATCH EXTRACTION RESULTS

ELEMENT	CONCENTRATIONS, μg/l WET FGD SLUDGE	CONCENTRATIONS, μg/l DRY FGD SLUDGE
SILVER	<60	—
ARSENIC	60-110	1-600
BARIUM	<400-1600	—
CADMIUM	<250	1-500
CHROMIUM	<100	—
MERCURY	<2	—
LEAD	<500	—
SELENIUM	30-100	20

(BASIS: 2 WET TESTS; 5 DRY)

COLUMN LEACHING TESTS

GENERAL OBSERVATIONS

- Data are limited
- Ions of soluble salts (TDS, Na⁺, Ca⁺⁺, SO₄⁼) initially follow washout behavior (CURVE 1) (e.g. initial TDS ~ 50,000 decreases to ~3000 mg/l at a given column liquid loading rate)
- Some constituents appear to have a delayed maximum leaching rate (e.g. Al, Cr, Zn: CURVE 2)
- The leaching rate of some trace constituents decreases readily so that the resulting concentrations in the column elutriate approach the detection limits (CURVE 3)

Figure 4. Summary of Physical-Chemical Characteristics of Power Plant Flue Gas Desulfurization Sludge*
*This figure is taken from EPRI Report EA-3236, Figure 7-2.

GEOHYDROCHEMICAL MODELS FOR SOLUTE MIGRATION—AN EVALUATION

An initial evaluation of existing geohydrochemical models and codes is being carried out by the researchers in Phase I. The first volume of the reports is in the final stages of preparation (Kincaid, Morrey, and Rogers, 1983). This volume contains descriptions of processes deemed important in solute migration and a selection, for initial evaluation, of 21 existing codes (Table 5) which deal with these processes. These codes employ one or more of the three components in geohydrochemical model(s): (1) hydrological flow and transport, (2) geochemical interactions influencing the availability of a constituent for migration, and (3) microbial transformation which may alter the constituent during migration. The second and third volumes are to be prepared by early 1984 and will report on the findings of simulations on the applicability of a subset of these selected codes for predicting solute migration around the utility industry disposal sites.

CHEMICAL ATTENUATION RATES, COEFFICIENTS, AND CONSTANTS IN LEACHATE MIGRATION

A two-volume report prepared under this Phase I research contract is being finalized for publication. In Volume I of the report (Rai and Zachara, 1983) data obtained from the available literature have been summarized for each of the elements. The quantitative results given allow thermodynamic data-based predictions of the relative stability of solid and aqueous species, observed or hypothesized solubility controls in the geochemical environments, and the rates of reactions for adsorption-desorption and precipitation-dissolution processes for chemical attenuation. Most of the data compiled pertains to single element and pure minerals.

Volume II of this report (Rai, Zachara, Schmidt, and Schwab, 1983) is an annotated bibliography of over 350 publications which appeared in existing literature prior to February 1983 and deals with research on chemical (precipitation-dissolution and adsorption-desorption) and biological (methylation and alkylation) attenuation mechanisms for Al, As, B, Ba, Be, Cd, Cl, Cr, Cu, F, Fe, Mn, Mo, Na, Ni, Pb, Sb, Se, SO_4, V, and Zn.

SUMMARY AND CONCLUSIONS

In this paper a chronological summary of the SWES project at EPRI is developed. Although the research is in an early stage of implementation, it is expected that thorough and deliberate planning has given the research a very high chance for successful completion. In the short-term (the next 3 to 5 years), the research results will be in the form of data which can be used in assessing the potential for release, transformation, and transport of inorganic solutes from utility industry solid residues. In the long term (the next 6 to 8 years), the research results are to provide predictive methods (mostly geohydrochemical model[s]) to quantitatively estimate the fate of solutes in the groundwaters.

TABLE 5. COMPUTER CODES SELECTED FOR EVALUATION IN SWES

	Code Designation	Description/Function
I. UNSATURATED FLOW AND TRANSPORT CODES:		
(1)	SESOIL	Compartmental
(2)	NRC-SLB	One-dimensional
(3)	OR-NATURE	One-dimensional
(4)	UNSAT1D	One-dimensional
(5)	FEMWATER/FEMWASTE	Multi-dimensional
(6)	TRUST/MLTRAN	Multi-dimensional
(7)	SATURN	Multi-dimensional
II. SATURATED FLOW AND TRANSPORT CODES:		
(8)	PATHS	Multi-dimensional-analytical
(9)	SWIP2/SWENT	Multi-dimensional
(10)	TRANS	One or two dimensional
(11)	VTT	One or two dimensional
(12)	FE3DGW	Two or three dimensional
(13)	USGS MOC	Two dimensional
(14)	AT123D	Multi-dimensional-analytical
III. EQUILIBRIUM GEOCHEMICAL CODES:		
(15)	GEOCHEM	Equilibrium, geochemistry, includes adsorption
(16)	MINTEQ	Equilibrium, geochemistry, includes adsorption
(17)	PHREEQE	Precipitation-dissolution
(18)	EQUILIB	Precipitation
(19)	EQ3/EQ6	Precipitation
IV. MICROBIOLOGICAL CODE:		
(20)	BIOFILM	Primarily organic compounds
V. HYDROLOGICAL AND GEOCHEMICAL CODE:		
(21)	FIESTA	One-dimensional flow and transport coupled with equilibrium geochemistry

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SESSION 10, PART I: DRY FGD: PILOT PLANT TEST RESULTS

Chairman: Theodore G. Brna
Industrial Environmental Research Laboratory
U.S. Environmental Protection Agency
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CURRENT STATUS OF DRY SO_2 CONTROL SYSTEMS

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ABSTRACT

This paper provides an update on commercial applications and research and development (R&D) activities involving three dry SO_2 control technologies: spray drying, dry injection, and electron-beam irradiation.

Spray drying continues to be the only commercially applied dry flue gas desulfurization (FGD) process, with two additional spray drying systems having been sold since mid-1982. To date, 17 commercial utility spray drying systems have been sold, totalling over 6,800 MWe. Six of these systems are currently operational with two of the systems having been turned over to the utility. In addition to the utility applications, there are 21 industrial spray drying units, 7 of which are operational. Several additional utility and industrial systems are expected to start up in the next 2 years. Demonstration- and pilot-scale testing of the spray dryer process is continuing with emphasis on high sulfur applications.

The first planned commercial application of dry injection technology has been announced for a 500 MWe utility. The recent demonstration-scale testing on a 22 MWe unit has been completed.

The electron-beam process is also in an early developmental state. Pilot-scale testing of the electron-beam/lime spray drying version of the process is scheduled to begin this fall.

INTRODUCTION

This paper updates the commercial and research and development (R&D) status of dry SO_2 control processes for utility and industrial boilers. Four EPA reports on the status of dry SO_2 control systems have been published to date (1, 2, 3, 4), the most recent in August 1983. A fifth update is currently being prepared, with completion scheduled for early 1984. The primary sources of information have been contacts with dry SO_2 control system vendors and purchasers and government and industry agencies involved with dry SO_2 control in the United States. Three types of dry SO_2 control (nonregenerable) processes will be covered in this paper, the latter one briefly because of coverage elsewhere during this symposium:

- Spray drying.
- Dry injection.
- Electron-beam irradiation.

Spray drying, which includes a fabric filter or an electrostatic precipitator (ESP) for collection of fly ash and desulfurization products, is the only one of the three processes to be commercially applied. Dry injection, a process involving injection of dry alkaline material directly into the flue gas and subsequent collection of the desulfurization products and fly ash in a fabric filter, is nearing commercial application. The electron-beam process, which involves reagent injection into the flue gas followed by electron-beam irradiation, is in an early developmental state.

A more detailed description of these processes, including some of the important process variables and design parameters, is presented below. The technology descriptions are followed by a discussion of the current commercial applications and R&D activities for each of the three processes.

TECHNOLOGY DESCRIPTIONS

SPRAY DRYING

Figure 1 is a schematic diagram of the spray drying FGD process. The gas containing fly ash and SO_2 enters the spray dryer and is contacted with a finely atomized alkaline slurry or solution. During the approximately 10-second residence time in the dryer, the flue gas is adiabatically humidified as the water in the slurry or solution is evaporated. Simultaneously, flue gas SO_2 reacts with the alkaline species to form solid sulfite and sulfate salts. The solids formed are dried to generally less than 1 percent free moisture. The flue gas, which has usually been humidified to within 11 to 28°C (20 to 50°F) of its adiabatic saturation temperature, passes through the dryer and into a downstream high efficiency particulate matter control device. In some designs a portion of the solids drop out of the dryer, but the bulk of the desulfurization products are collected with fly ash in a fabric filter or an ESP. The reaction between the alkaline material and flue gas SO_2 continues as the gas passes through the ductwork and the fabric filter or ESP.

Table 1 lists the major design and operating factors for spray drying FGD. These considerations are discussed in more detail below.

Using a spray dryer as a flue gas contactor involves adiabatically humidifying the flue gas to within a specified number of degrees above saturation. With set conditions for inlet flue gas temperature and humidity and for a specified approach to saturation temperature, the amount of water which can be evaporated into the flue gas is set by energy balance considerations. Liquid-to-gas ratios are generally in the range of 0.27 to

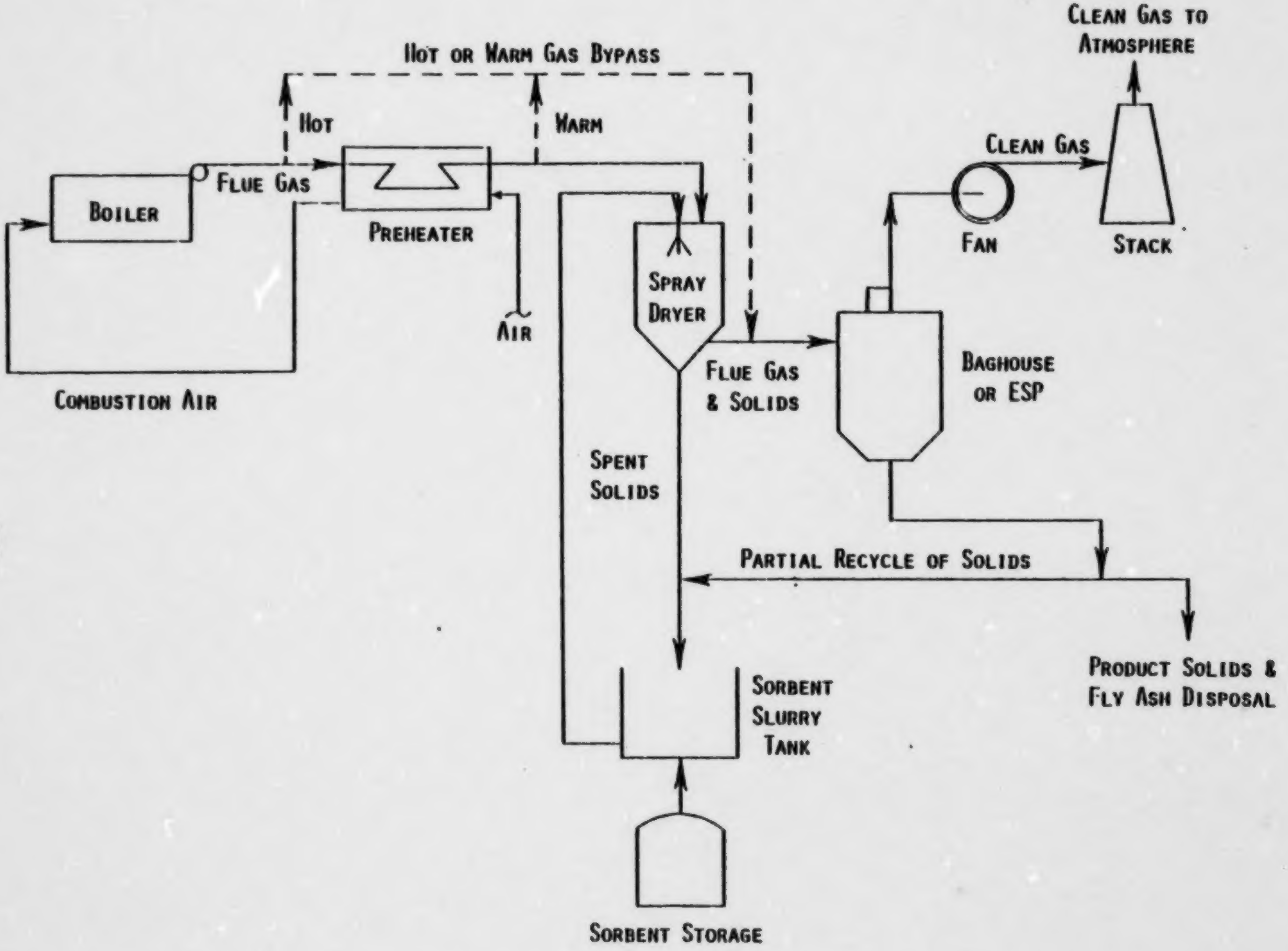

Figure 1. Typical spray dryer/particulate collection flow diagram.

TABLE 1. IMPORTANT DESIGN AND OPERATIONAL FACTORS FOR SPRAY DRYING FGD SYSTEMS

Operational Factors

- Approach to saturation at dryer outlet
- Sorbent stoichiometry
- Inlet SO_2 concentration
- Temperature drop over the spray dryer
- Fly ash alkalinity
- Gas residence time

Design Factors

- Spray dryer design (gas distribution, gas exit configuration)
- Atomizer type
- Slaker type
- Use of solids recycle
- Use of flue gas bypass

0.40 liter/m^3 (0.2 to 0.3 gal./kcf). The sorbent stoichiometry is varied by raising or lowering the concentration of a solution or weight percent solids of a slurry containing this set amount of water. While holding other parameters such as temperature constant, the obvious way to increase SO_2 removal is to increase sorbent stoichiometry. However, as sorbent stoichiometry is increased to raise the level of SO_2 removal, two limiting factors are approached:

- Sorbent utilization decreases, raising sorbent and disposal costs.
- An upper limit is reached for the solubility of the sorbent in the solution, or for the weight percent of sorbent solids in a slurry.

There are at least two methods of circumventing these limitations. One method is sorbent recycle, using the solids that have either dropped out in the spray dryer or collected in the particulate emission control device. Recycle increases sorbent utilization and can also increase utilization of any alkalinity in the fly ash.

The second method of avoiding the above limitations on SO_2 removal is to operate the spray dryer at a lower outlet temperature; that is, a closer approach to saturation. Operating the spray dryer at a closer approach to saturation has the effect of increasing both the residence time of the liquid droplets and the residual moisture level in the dried solids. As the approach to saturation decreases, SO_2 removal rates and sorbent utilization generally increase dramatically.

The approach to saturation at the spray dryer outlet is set by either the requirement for a margin of safety to avoid condensation in downstream equipment or restrictions on stack temperature. The spray dryer outlet can be operated at temperatures lower than these restrictions would otherwise allow if some warm or hot gas is bypassed around the spray dryer to reheat the dryer outlet gas. Warm gas (from downstream of the boiler air heater) entails no energy penalty, but bypassing untreated gas for reheating can limit overall SO_2 removal efficiencies. Significantly less hot gas (upstream of the air heater) is required for heating, but using hot gas decreases the energy available for air preheating. Figure 1 illustrates these two reheat options.

Several similarities exist between the designs of the commercial spray drying FGD systems sold to date. All of the systems except one use lime as the sorbent; the exception uses commercial soda ash. In addition, all but two of the purchased systems include a fabric filter for collection of fly ash and waste products. Two utility systems will have an ESP instead of a fabric filter. Reverse-air fabric filters have been selected for utility systems, while pulse-jet units have been the choice for most industrial systems.

Most commercial system designs use rotary atomizers, although nozzle atomizers are used in some designs. Solids recycle and flue gas bypass for limited reheat are site-specific design options. Recycle, however, is being included more often than not, primarily because it improves sorbent utilization and lowers reagent costs. The use of recycle and flue gas bypass depends primarily on the SO_2 removal requirements.

Several differences between the commercialized designs of spray dryer FGD systems are evident. The differences can be categorized into four major areas:

- Spray dryer design and operation.
- Atomization.
- Sorbent preparation (lime slaking).
- Fabric filter design and operation.

Variations in spray dryer design include the shape of the dryer, gas disperser configurations, multiple or single rotary (or nozzle) atomizers per dryer, one- versus two-point flue gas admission to the spray dryer, and solids collection from the spray dryer. Each vendor claims his design to be suitable for FGD and to have certain advantages. Some important considerations in evaluating the various designs include:

- Liquid/gas contact.
- Turndown capability.
- Potential for plugging or carryover of wet solids to baghouse during upset conditions.

Differences in spray dryer operation include gas residence time and the approach to saturation at the dryer outlet. Gas residence times range from 7 to 12 seconds in most commercial designs; most operate in the 10- to 12-second range. Many commercial system designs call for a relatively close approach to saturation at the dryer outlet [10 - 14°C (18 - 25°F)]. These close approaches are common where SO_2 removal efficiency requirements approach 85 to 90 percent, but they may also be used to decrease reagent use in lower removal efficiency applications. Some designs, however, are based on a wider approach to saturation [17 - 28°C (30 - 50°F)].

Atomization techniques vary with regard to the use of rotary or nozzle atomizers, atomizer wheel speed in rotary atomizers, atomizing fluid (air or steam) for nozzles, internal or external mixing of fluids in nozzles, and the number of atomizers. Rotary atomizers may be used on a single or multiple atomizer per dryer basis. Several commercial utility systems use three atomizers per dryer, but most sold to date have a single atomizer per dryer. Rotary atomizers are generally thought to have certain advantages over nozzle atomizers. Specifically, unlike nozzle atomization, the size of the droplets produced by rotary atomizers is independent of slurry flow rate, providing for good turndown capability. Some plugging problems have occurred in nozzles using steam for atomization. One potential advantage of nozzle atomizers lies in the fact that they are mechanically simpler than rotary atomizers.

The method used to slake the lime also varies for commercial system designs. Ball mill, paste, and detention slakers are all used, but ball mill and paste slakers are more common. Each method is perceived to have certain advantages. Ball mill slakers generally produce a more finely ground and possibly more reactive slurry and, unlike paste slakers, no grit removal and disposal are required. Paste slakers, on the other hand, have lower capital costs, lower noise levels, and lower power requirements. Also, they generally produce a less abrasive slurry than do ball mills, which could reduce the potential for wear and maintenance of system pumps and piping.

As stated above, fabric filters have been chosen over ESPs in nearly all the commercial spray drying applications to date. Fabric filters may have an advantage over ESPs in that unreacted alkalinity in the solids and fly ash collected on the fabric surface can react with remaining SO_2 in the flue gas as the gas passes through the fabric filter. Some process developers have reported that SO_2 removal across the fabric filter can account for at least 10 percent of the total system SO_2 removal. In some applications, however, an ESP might be more desirable than a fabric filter. The factors that are important in making the choice between ESPs and fabric filters include:

- Use of recycle (increases dust loading to ESP, thus increasing ESP size and cost).
- Fly ash resistivity (high ash resistivity often requires larger, more expensive ESPs).
- Pressure drop (an ESP has a lower pressure drop and related costs than a fabric filter).

Fabric filter designs for spray dryer FGD systems vary primarily with regard to bag fabric, cleaning frequency, and cleaning mode. Other more subtle variations, too numerous to mention, between the well-established fabric filter design practices of the different vendors also exist. Teflon-coated fiberglass bags have been selected for many systems, although acrylic bags may be a lower cost alternative for low gas temperatures.

DRY INJECTION

A generalized flow diagram of the dry injection process is shown in Figure 2. The process involves pneumatic injection of a dry, powdery sodium compound into the flue gas with subsequent particulate collection in a fabric filter. The point of alkali injection may vary from upstream of the air preheater to the inlet of the fabric filter. Reaction between the reagent and SO_2 occurs both in the duct and on the filter bag surface (5,6).

Although other alkaline reagents (e.g., lime and limestone) have been tested, only certain sodium compounds have shown the capability for high SO_2 removal from the flue gas. Nahcolite and trona ores, which contain naturally occuring sodium compounds, appear to be the most promising reagents for dry injection in terms of reactivity and cost. Nahcolite, which is usually associated with western oil shale reserves, contains about

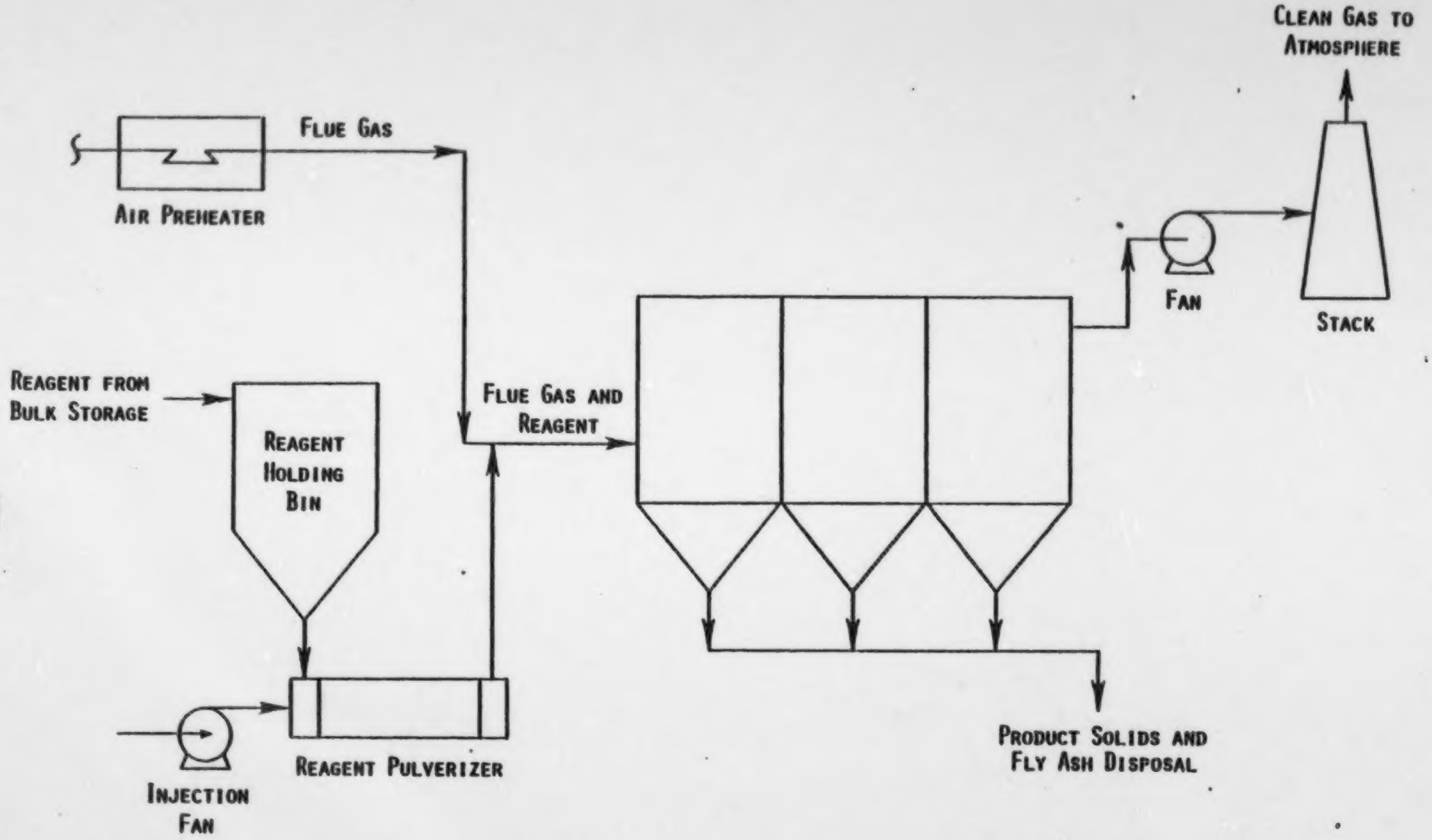

Figure 2. Dry alkali injection flow diagram.

80 percent sodium bicarbonate ($NaHCO_3$). Trona ore, which is found in large deposits in Wyoming and California, contains sodium carbonate (Na_2CO_3) and sodium sesquicarbonate ($Na_2CO_3 \cdot NaHCO_3 \cdot 2H_2O$). Nahcolite has been shown to be somewhat more reactive with SO_2 in the flue gas than trona (5,6).

Major factors that affect SO_2 removal by nahcolite and trona injection include stoichiometric ratio and the flue gas temperature at the point of injection. Higher SO_2 removals are obtained at higher normalized stoichiometric ratios (equivalent moles Na_2O per mole of inlet SO_2). However, higher stoichiometric ratios also result in lower reagent utilization (5). Injection of the reagent at too low a temperature will reduce the initial SO_2 reaction rate and may limit overall SO_2 removal. For nahcolite it appears that SO_2 removal drops off dramatically at injection temperatures below 135°C (275°F) (6,7).

Other parameters that are important in the dry injection process include reagent particle size, mode of injection (batch, semi-batch, or continuous), baghouse air-to-cloth ratio, and bag cleaning frequency.

ELECTRON-BEAM PROCESS

The electron-beam (E-beam) process involves the irradiation of flue gas containing a reactant, such as ammonia or lime. The process removes both SO_2 and NO_x from the flue gas and results in a dry waste product.

A schematic diagram of the E-beam/ammonia process is shown in Figure 3. In this process, flue gas from a fly ash collection device is cooled and humidified in a quench tower. The resulting gas moisture content is about 10 percent. Ammonia is injected into the cooled gas and the gas is passed through an E-beam reactor. In the reactor, oxygen and water are excited and ionized to form the radicals [OH], [O], and [HO_2] by the application of electrons at a dose of 1 to 3 Mrads (1 Mrad is equivalent to 10 joules/g of flue gas). These radicals react with SO_2 and NO_x to form sulfuric acid (H_2SO_4) and nitric acid (HNO_3). The acids are neutralized by ammonia and water in the flue gas to form solid ammonium sulfate (NH_4SO_4) and ammonium sulfate nitrate [$(NH_4)_2 \cdot SO_4 \cdot NH_4NO_3$]. The reaction time for formation of the sulfate and nitrate salts is less than 1 second. Product solids are collected in a hopper below the E-beam reactor or in a downstream particulate collector.

In another version of the E-beam process, the water quench tower is replaced with a lime-based spray dryer. SO_2 is removed from the flue gas in the spray dryer and NO_x and additional SO_2 are removed in the E-beam reactor. Reactions in the E-beam reactor to form sulfuric and nitric acids occur in the same manner as in the ammonia process. The acids are neutralized by Ca(OH) to form calcium salts ($CaSO_4$, $Ca(NO_3)_2$, and $CaSO_3$).

Bench- and pilot-scale studies of E-beam processes have shown that variables important to performance include gas moisture content, gas temperature, oxygen content, reagent ratio, and electron dosage. In addition, efficient penetration of the gas stream by the beam requires a unique discharge pattern and other special design considerations.

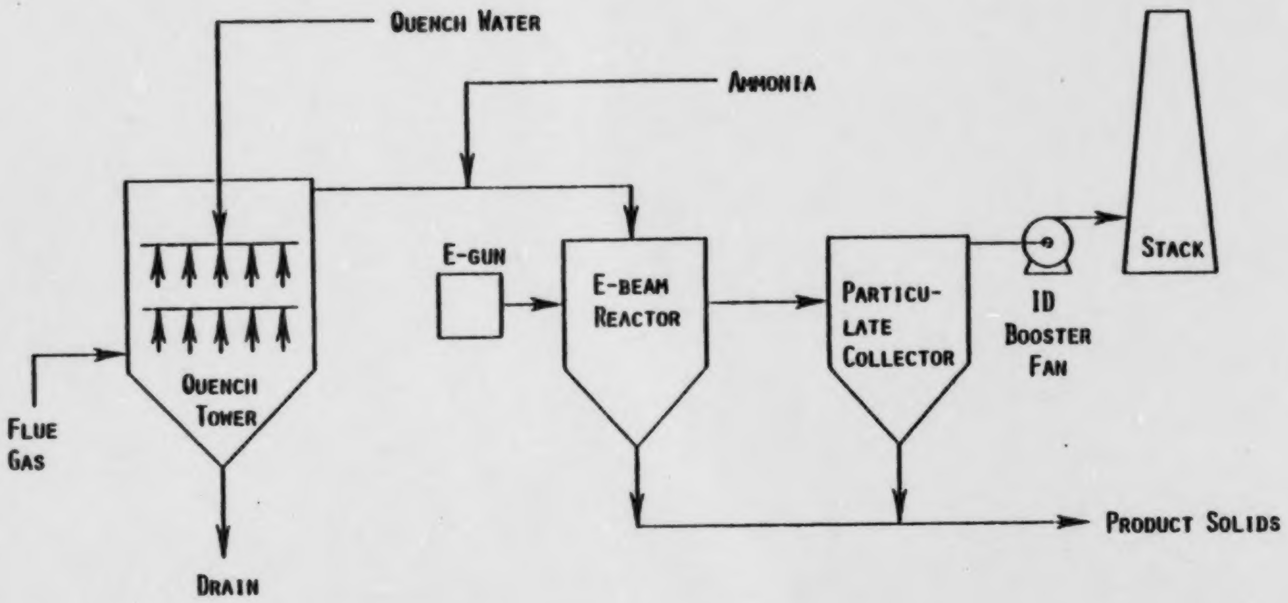

Figure 3. E-beam/ammonia process flow diagram.

DEVELOPMENT STATUS

Of the three dry SO_2 control technologies considered here, only spray drying has been commercially applied. However, dry injection technology is nearing commercialization with plans for a 500 MWe system. Research and development activities are being conducted for all three technologies by vendors, government agencies, and industry organizations.

SPRAY DRYING

Through October 1983 there have been 17 commercial utility systems sales, totaling over 6,800 MWe. The general characteristics of the utility systems, including vendor, size, location, design criteria, and projected start-up date or current operational status, are shown in Table 2. The commercial utility spray drying systems have been selected for units firing relatively low sulfur western coal or lignite (generally less than 1.5 percent sulfur fuel). SO_2 removal guarantees range from 60 to 91 percent, with design controlled emission levels ranging from 77 to 516 ng/J (0.18 to 1.2 lb/10^6 Btu). Some of the guarantees stem from the need to meet the most recent New Source Performance Standards (NSPS) for utility boilers (8).

Two commercial utility spray drying systems have been sold since mid-1982, and a third system was awarded but is currently on hold. These awards were all to Joy Industrial Equipment Company and Niro Atomizer, Inc. The recently awarded utility systems are for units firing coal/lignite with average sulfur contents of less than 1 percent.

According to the PEDCo survey prepared for EPA, 24 utilities considering FGD for new coal-fired boilers have not yet made awards (9). Of these utilities, two are considering only dry FGD, while nine utilities did not specify whether they were considering only wet, only dry, or both. Table 3 shows the utilities, the planned units, and reported coal sulfur contents for which FGD is being considered. The data in Table 3 indicate that dry FGD is not yet being widely considered for utility applications involving moderate or high sulfur coal.

The six commercial utility spray drying systems now operational are listed in Table 2. Of these the sodium-based system at Coyote Station was turned over to Montana-Dakota Utilities for operation following performance tests conducted in August 1982 (10). Earlier test results reported for this system (11) showed 70 percent SO_2 removal at a soda ash utilization of 75 percent and an approach to saturation of 36°C (65°F).

The 110 MWe unit at Northern States Power Company's Riverside Station, initially a demonstration spray dryer coupled with existing ESPs and now coupled to a fabric filter treating flue gas from two boilers, has also been turned over to the utility. Process testing of this system was performed during the summer of 1983 under EPA and EPRI funding. Results from these tests are scheduled to be presented later at this symposium.

TABLE 2. COMMERCIAL UTILITY SPRAY DRYING SYSTEMS

System Purchaser	Station/Location	Size[a] MWe	Start-up Date/Status	Coal	SO_2 Removal Guarantee	Outlet SO_2 ng/J (lb/10^6 Btu)[b]	Vendor[c]
Otter Tail Power Co. (Operator: Montana-Dakota Utilities)	Coyote, Unit 1 (Beulah, ND)	440	Operational. Turned over to utility.	North Dakota lignite; 0.78% S average.	70% for all coal.	516 (1.2)	Rockwell/ Wheelabrator-Frye
United Power Assoc.	Stanton, Unit 1A (Stanton, ND)	60	Operational. Not turned over to utility.	North Dakota lignite; 0.77% S average; 1.94% S maximum.	91% for max. S coal.	258 (0.6)	Cottrell Environmental/ Komline-Sanderson
Marquette Board of Light and Power	Shiras, Unit 3 (Marquette, MI)	44	Operational. Not turned over to utility.	Western subbituminous; 1.5% S maximum.	80% design efficiency.	404 (0.94)	G.E. Environmental Services
Basin Electric Power Coop.	Laramie River, Unit 3 (Wheatland, WY)	575	Operational. Not turned over to utility.	Wyoming subbituminous 0.54% S average; 0.81% S maximum.	82% for avg. S coal; 90% for max. S coal;	86 (0.20)	Babcock and Wilcox
Colorado Ute Electric Assoc.	Craig, Unit 3 (Craig, CO)	447	Late 1983. Under construction.	Western subbituminous 0.7% S design coal.	87% for design coal.	86 (0.20)	Babcock and Wilcox
Basin Electric Power Coop.	Antelope Valley, Unit 1 (Beulah, ND)	440	Operational. Not turned over to utility.	North Dakota lignite; 0.68% S average; 1.22% S maximum.	62% for avg. S coal; 78% for max. S coal.	335 (0.78)	Joy/Niro
Basin Electric Power Coop.	Antelope Valley, Unit 2 (Beulah, ND)	440	April 1985.	North Dakota lignite; 0.68% S average; 1.22% S maximum.	89% for max. S coal.	168 (0.39)	Joy/Niro
Tucson Electric Power	Springerville, Unit 1 (Springerville, AZ)	370	June 1985.	New Mexico subbituminous; 0.69% S average.	61%.	258 (0.6)	Joy/Niro
Tucson Electric Power	Springerville, Unit 2 (Springerville, AZ)	370	June 1987.	New Mexico subbituminous; 0.69% S average.	61%.	258 (0.6)	Joy/Niro
Platte River Power Authority	Rawhide, Unit 1 (Fort Collins, CO)	280	December 1983. Under construction.	Western subbituminous 0.3% S average; 0.44% S maximum.	80%.	77 (0.18)	Joy/Niro

(continued)

TABLE 2. (CONTINUED)

System Purchaser	Station/Location	Size[a] MWe	Start-up Date/Status	Coal	SO_2 Removal Guarantee	Outlet SO_2 ng/J (lb/10^6 Btu)[b]	Vendor[c]
Sunflower Electric	Holcomb, Unit 1 (Holcomb, KS)	319	In start-up	Wyoming subbituminous; 0.34% S average.	80%.	116 (0.27)	Joy/Niro
Sierra Pacific Power	North Valmy (Valmy, NV)	270	1984.	Western subbituminous; 0.4 to 1.0% S.	76% for all coal.	258 (0.6)	Rockwell
Grand River Dam Authority-State of Oklahoma	Pryor, Unit 2 (Pryor, OK)	520	March 1985.	Western subbituminous; 0.4 to 1.5% S.	85% for all coal.	NR	Flakt
Northern States Power Company	Sherburne County Unit 3 (Becker, MN)	860	1990.	Sarpy Creek subbituminous; 0.9% S average; 0.4 – 2.3% S range.	70% minimum.	258 (0.6)	Joy/Niro
Cajun Electric	Oxbow, Unit 1 (Coushatta, LA)	563	Projected for 1989 but need for system being re-evaluated	Louisiana lignite; 0.93% S average.	80%.	258 (0.6)	Joy/Niro
Central and Southwestern Services	Coleto Creek Unit 2 (Corpus Christi, TX)	710	1988.	Subbituminous; 0.4% S average.	70%.	NR	Joy/Niro
Northern States Power Company	Riverside, Units 6 and 7 (Minneapolis, MN)	110	Operational. Turned over to utility.	Varies with test series.	Varies with test program.	516 (1.2)	Joy/Niro
Pacific Power and Light	Jim Bridger, Unit 2, (Rock Springs, WY)	100	Shut down. Demonstration testing by vendor completed. No plans for restart.	Western subbituminous; 0.56% S average; 0.8% S maximum.	Lime: 62-75% Soda ash: 74-86%	129 (0.3)	Flakt

[a] Gross MWe output.

[b] Source for SO_2 outlet emissions: Ireland, P. A. (Sterns-Roger) Status of Spray Dryer Flue Gas Desulfurization. EPRI CS-2209. Final Report. EPRI, Palo Alto, CA., January 1982, p. 2-3.

[c] Rockwell/Wheelabrator-Frye is no longer a joint venture. Both are, however, selling spray drying systems. Joy/Niro: Joy takes the lead in utility sales, while Niro takes the lead in industrial sales. G.E. Environmental Services was formerly Buell Emission Control Division, Envirotech Corp. Spray dryer supplied by Anhydro A/S and fabric filter supplied by G.E.E.S. Cottrell Environmental Sciences supplies fabric filter and Komline-Sanderson supplies spray dryer.

TABLE 3. UTILITIES PLANNING OR CONSIDERING FGD SYSTEM THAT HAVE NOT AWARDED BIDS[a] (as of August 3, 1983)

Utility	Type of FGD System Considered	Coal (%S)	Projected Unit Start-Up
Atlantic City Electric	Wet	3.25	1988
Buckeye Power	Wet	NR[b]	1990
Central Illinois Light	Wet	3.3	1989
Central Maine Power	Wet	2.23	1989
Cincinnati G&E	NR	4.0	1988
Colorado Ute Electric	NR	0.5	1988
Delmarva Power & Light	Wet	2.5	1987
Desert Gen & Trans	Wet	0.5	1992
Florida Power & Light	NR	NR	2 units; 1992, 1993
General Public Utilities	NR	3.5	2 units; 1993, 1994
Indianapolis Power & Light	Wet	3.5	3 units; 1990
Kentucky Utilities	Wet	3.5	1994
Louisville G&E	Wet	4.0	2 units; 1986, 1990
Nebraska Public Power	Dry	0.36	1987
Nevada Power	NR	NR	4 units; 1984-87
Orlando Utilities Commission	Wet	NR	1987
Pacific Power & Light	Wet (retrofit)	0.56	3 units; 1986-90
	Dry	NR	1986
Salt River Project	Wet	0.6	1989
Seminole Electric	NR	NR	2 units; 1988, 1990
Southwestern Electric Power	NR	1.5	2 units; 1990
Texas Utilities	NR	0.8 (lignite)	1989
WA Water & Power	Wet	NR	4 units; 1990-94
West Texas Utilities	NR	0.34	1996

[a] Source: Reference 9.

[b] NR - Not reported.

Three of the other four operational spray drying systems (Stanton, Shiras, and Antelope Valley) have all recently met performance guarantees. However, no test results have been reported yet for any of these systems.

One other utility system is in the initial start-up stage, and two systems are scheduled for start-up before the end of 1983.

In addition to the utility systems, a number of industrial coal-fired boilers are, or will be, equipped with spray drying FGD systems. Table 4 gives a summary of the 7 operational and 14 planned industrial spray drying units. Coal sulfur contents for the industrial applications range from 0.6 to 3.5 percent, with seven systems designed for coal sulfur contents of 3 percent or greater. Removal guarantees for the industrial systems range from 70 to 90 percent.

Performance test results have been reported for five of the operational industrial systems. A continuous monitoring program conducted by EPA on the Celanese system showed 70 percent average SO_2 removal over a 23-day test period (12). Removal averaged 75 percent on days when the daily inlet SO_2 concentrations were 1720 ng/J (4.0 lb/10^6 Btu) or greater. It should be noted that the tests were conducted during a period of fluctuating boiler operation.

Performance tests of the Strathmore Paper Company system showed an average SO_2 removal of 92.4 percent (90.1 to 96.7 percent range) at an inlet SO_2 concentration of about 2000 ppm (13). Data on lime stoichiometric ratio or other operating conditions required to achieve these removals were not included in the performance test results. This system and the Celanese unit, which do not employ recycling of solids, have now been turned over to the system purchasers.

Compliance tests conducted on the Argonne system showed SO_2 removals ranging from 79.7 to 95.6 percent at lime stoichiometries ranging from 0.8 to 2.0, respectively (14). These tests were conducted on a 3-percent sulfur coal at an approach to saturation of approximately 13°C (23°F). In this system, receiving flue gas from a stoker-fired boiler and described in a presentation to be made later in this session, solids from the flue gas cleaning system are recirculated to the spray dryer.

This system has now been turned over to the Argonne National Laboratory, and EPA-funded process testing of the system is on-going. Preliminary results from a 100-hour test run on a 4.6 percent sulfur coal showed 90 percent SO_2 removal at a lime stoichiometry of 1.4.

The Container Corporation and Austell Box Board systems have both met compliance requirements, and the former system has been turned over to the system purchaser. Compliance tests on the Container Corporation system showed average outlet SO_2 emissions of 14.3 ng/J (0.03 lb/10^6 Btu) on a 0.6 percent sulfur coal (15). Tests on the Austell Box Board system (16) showed an average outlet SO_2 emission rate of 366 ng/J (0.85 lb/10^6 Btu). No data on reagent ratios have been reported for these two systems. The system at GM-Buick Motors and one unit at the University of Minnesota have only recently become operational, and no performance data have yet been released.

TABLE 4. COMMERCIAL INDUSTRIAL BOILER SPRAY DRYING SYSTEMS

System Purchaser	Location	Size, Mg/hr (lb/hr) Steam	Start-up Date/Status	Coal	SO_2 Removal Guarantee	Vendor[a]
Celanese Fibers Co.	Cumberland, MD	50 (110,000)	Operational. Turned over to purchaser.	1.5% S to 2.5% S eastern coals.	70% for 1.5% S coal. 87% for 2.5% S coal.	Rockwell/ Wheelabrator-Frye
Strathmore Paper Co.	Woronco, MA	39 (85,000)	Operational. Turned over to purchaser.	2.3 to 3% S eastern coal.	75%.	Mikropul
University of Minnesota	Minneapolis, MN	Two units @ 40 MW each	One unit operational. Second start-up in September 1983. Not turned over to purchaser.	Subbituminous coal; 0.6 to 0.7% S.	70%.	Flakt
Argonne National Lab	Argonne, IL	77 (170,000)	Operational. Turned over to purchaser.	Bituminous coal; 3.5% S.	78.8% (516 ng/J or 1.2 lb/10^6 Btu outlet SO_2).	Niro/Joy
Container Corporation	Philadelphia, PA	77 (170,000)	Operational. Turned over to purchaser.	1.0% S eastern coal.	90%.	Ecolaire
GM-Buick Motors	Flint, MI	204 (450,000)	Operational. Not turned over to purchaser.	Indiana bituminous coal; 1 to 3% S.	70 to 90%.	Niro/Joy
Fairchild AFB	Units 1,2,3 Spokane, WA	3 @ 50 (110,000)	Initial start-up stages.	1.0% S western coal.	85%.	Niro/Joy
Austell Box Board Co.	Austell, GA	114 (250,000)	Operational. Not turned over to purchaser.	Bituminous coal; 1.0 to 2.5% S.	516 ng SO_2/J (1.2 lb SO_2/10^6 Btu outlet).	Wheelabrator-Frye
Puget Sound Naval Shipyard	Units 1,2,3 Bremerton, WA	3 @ 64 (140,000)	Late 1987.	1.6% S maximum.	84%.	G.E. Environmental Services
Maelstrom AFB	Units 1,2,3 Great Falls, MT	3 @ 41 (90,000) hot water	Spring 1985.	Western subbitiuminous coal; 1.0% S.	85%.	Niro/Joy
Griffis AFB	Units 1,2,3,4 Rome, NY	4 @ 41 (90,000)	Late 1984	Bituminous; 3% S	85% (0.71 lb SO_2/10^6 Btu outlet).	Ecolaire

[a] Rockwell/Wheelabrator-Frye is no longer a joint venture. Both are, however, selling spray drying systems. Niro of Joy/Niro takes the lead in industrial system sales, and Joy takes the lead in utility system sales.

In addition to the commercial sales, there are a number of spray drying research and development activities currently underway. Table 5 lists the major current activities. These programs, which are being funded by EPA, EPRI, and DOE, are designed to address information needs in the areas of high sulfur coal applications, alternate spray drying sorbents, and sustained operation at a close approach to saturation. Results from several of the programs are scheduled to be presented at this symposium. On-going research by several vendors stresses the further improvement of atomization techniques with goals of longer service life, lower energy consumption, and improved slurry droplet size distribution. Rotary and nozzle atomizers are being compared on the basis of cost and performance.

DRY INJECTION

The first planned commercial application of dry injection technology has been announced by Public Service Company of Colorado for a 500 MWe unit scheduled for start-up in 1990 (17). The system, which will use trona ore as the sorbent, will be designed for 70 percent SO_2 removal on a 0.4 percent sulfur western coal. The trona will be injected just upstream of the fabric filter into flue gas at approximately 132 to 138°C (270 to 280°F). A clay- and plastic-lined landfill will be used for solids disposal.

The 22 MWe demonstration testing at Public Service of Colorado's Cameo Station has been completed. Final reports on the EPRI-sponsored dry injection program have been prepared (18,19). Results of the testing showed 70 percent SO_2 removals for trona and nahcolite at stoichiometric ratios of approximately 1.3 and 0.8, respectively. These tests were conducted on a low sulfur western coal yielding flue gas at 450 ppm SO_2.

A recent cost study based on the Cameo results indicates that dry injection may be less expensive than spray drying, depending on reagent cost, coal sulfur content, and SO_2 removal requirements (7). Estimated costs for nahcolite and trona dry injection systems and a lime spray drying system are shown in Table 6 for two-500 MWe units firing 0.48 percent sulfur coal with 70 percent SO_2 removal. These estimates assume nahcolite (70 percent $NaHCO_3$) and trona (85 percent Na_2CO_3) costs of $100 and $75 per ton, respectively, and sodium-based waste disposal costs of $7.40/ton.

The application of trona dry injection had been previously constrained by questions regarding SO_2 removal limitations and cost. However, the recent demonstration-scale studies have shown that SO_2 removal efficiencies of 70 to 80 percent can be achieved with trona on low sulfur coals at reasonable stoichiometric ratios. Trona ore is currently mined in large quantities for conversion to sodium carbonate.

The commercial application of nahcolite dry injection hinges on the economic availability of nahcolite. Nahcolite is not currently mined in the United States, but at least one firm has announced intentions to develop a nahcolite mining operation and several other companies are investigating the possibility of supplying nahcolite through solution mining techniques (20,21).

TABLE 5. MAJOR CURRENT SPRAY DRYING R&D ACTIVITIES

Vendor, Agency, or Industry Group	Size	Location	Comments
Electric Power Research Institute (EPRI)[a]	255 m^3/min[b] (9000 cfm)	Arapahoe pilot plant; Denver, CO.	Test program began in March 1982. Testing planned through December 1983.
EPRI/EPA[a]	100 MWe	Riverside Station, Northern States Power. Minneapolis, MN.	Ten weeks of testing beginning in July 1983, including 3.4% S coal.
EPA	77 Mg/hr (170,000 lb/hr) steam	Argonne National Lab. Argonne, IL.	Performance test (90-day), starting in November 1983, on 3.4% S coal.
DOE/Babcock and Wilcox[a]	25 m^3/min (900 scfm)	B&W Alliance Laboratory. Alliance, OH.	Two-year test program began in mid-1982 on eastern high sulfur coals.

[a] Papers on these studies will be presented at this symposium.

[b] Actual flue gas conditions.

TABLE 6. COMPARISON OF COSTS FOR NAHCOLITE AND TRONA DRY INJECTION AND LIME SPRAY DRYING ON TWO NEW 500-MW COAL-FIRED POWER-GENERATING UNITS

FGD Process	Capital Cost ($/kW)	Levelized Costs (mills/kWh)
Trona injection	89.1	8.6
Nahcolite injection	90.4	9.6
Lime spray drying	175.9	10.8

Notes: All costs include baghouse for particulate control. Upper midwest plant firing 0.48 percent sulfur coal at 8,020 Btu/lb. 1983 revenue requirements.

Trona system: 70 percent SO_2 removal, $75/ton for trona (85 percent Na_2CO_3), $7.40/ton for waste disposal.

Nahcolite system: 70 percent SO_2 removal, $100/ton for nahcolite (70 percent $NaHCO_3$), $7.40/ton for waste disposal.

Lime system: 70 percent SO_2 removal, $60/ton for lime, $4.60 for waste disposal.

Source: Reference 7.

Another important consideration for dry injection applications is the high solubility and leachability of the sodium-based wastes. Application of dry injection may be limited in some locations by the costs of disposing of the wastes in an environmentally safe manner.

ELECTRON-BEAM PROCESS

The electron-beam process is in an early developmental state. The DOE has signed cost-sharing agreements with Research-Cottrell and EBARA/Avco-Everett to conduct pilot-scale demonstrations of E-Beam processes. Research-Cottrell will be developing the E-beam/lime process and EBARA/Avco-Everett will be developing the E-beam/ammonia process.

Testing of the E-beam/lime process is scheduled to begin in late 1983 on a 283 to 425 m^3/min (10,000 to 15,000 acfm) pilot plant at TVA's Shawnee power plant. The tests will be conducted on flue gas from firing a 2.5 to 3 percent sulfur coal (22). Test plans for a 570 Nm^3/min (20,000 scfm) pilot E-beam/ammonia system are being finalized with plans calling for construction to begin in October 1983 (23). This unit will also be tested on flue gas from firing a 2.5 to 3 percent sulfur coal.

SUMMARY

Currently over 6,800 MWe of electrical utility generating capacity has been committed to spray drying SO_2 control. Utilities are operating two systems (550 MWe), and four systems (1,120 MWe) are operational but not yet turned over for utility operation. Recent sales of industrial spray drying have been mainly to Federal government installations, with the industrial systems now sold totaling 21 units. Seven of these units are operational, with four units now being operated by their owners. While spray drying currently remains as the only commercially applied dry FGD technology, plans for a new 500 MWe electrical generating unit specify the dry injection of trona for SO_2 control at a Colorado site.

Research and development (R&D) on spray drying and the E-beam process is continuing. Both EPA and EPRI have active spray drying R&D programs. DOE is supporting the E-beam studies which are currently at the pilot level.

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ACID RAIN PREVENTION THRU NEW SO_x/NO_x DRY SCRUBBING PROCESS

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ABSTRACT

The issue of acid rain has attracted much attention both in the U.S. and in Europe. To cope with problems associated with acid rain Niro Atomizer is developing a dry scrubbing process for simultaneous removal of SO_x and NO_x.

A description of the process is given in this paper. The equipment used is essentially the same as used in the over 5000 MW utility dry scrubbers currently in operation, start-up or under construction.

Pilot plant test results from the Copenhagen dry FGD facility and results achieved during a full-scale demonstration of the process at one of JOY/NIRO's operating dry scrubbers are presented.

Waste product characteristics are shown and compared with EPA standards.

Finally, the process economics are analyzed and compared with other existing processes for SO_x/NO_x removal.

INTRODUCTION

In recent years the issue of acid rain has attracted much attention both in the United States and in Europe.

In the U.S.A. signs of acidification of the environment have become obvious in the North-East. Discussions focus on how big a reduction of SO_x and NO_x should be required (1). The U.S. EPA is evaluating which technique or combination of techniques would be the overall most cost effective in meeting the possible requirements (2, 3).

In Europe the 1982 Stockholm conference "Acidification Today and Tomorrow" (4) likewise concluded that dispersion of SO_x and NO_x arising from combustion of fossil fuels has a dramatic impact on the environment. Early this year a 30% reduction of SO_x

was suggested by Scandinavian countries, and recently the German conference "Acid Deposition - A Challenge for Europe" (5) again focused on the effects of acid deposition and abatement technologies available.

A number of techniques has been developed for reduction of SO_x and NO_x separately. Recent developments on low NO_x burner technology look promising. Considerable progress has been made in this area and achievable emission levels are approaching the 0.1 - 0.2 lb/MBtu range (6).

To encourage development of processes for simultaneous removal of SO_x and NO_x the U.S. Department of Energy and the U.S. EPA have funded a number of research programs that should lead to more cost effective methods for SO_x/NO_x control (7).

Dry scrubbing has in the recent years gained acceptance as an attractive FGD technique for many utility companies. Niro Atomizer has developed a new dry scrubbing process for simultaneous removal of SO_x and NO_x (8). The results of this development are presented in this paper.

DESCRIPTION OF NEW SO_x/NO_x PROCESS

The new dry scrubber process for simultaneous SO_x and NO_x removal (9) utilizes the same equipment as currently supplied for the JOY/NIRO dry FGD system. Normal operating conditions in these systems are low approach temperatures such as 18°F to achieve the lowest lime consumption and the highest SO_x scrubbing efficiency. Under these conditions virtually no NO_x absorption is achieved.

The new SO_x/NO_x process is schematically shown in Figure 1.

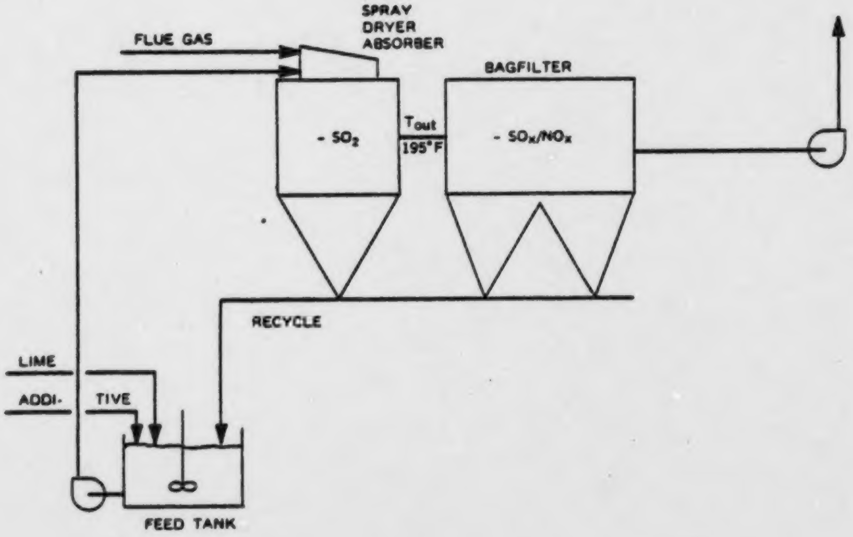

Figure 1. Niro Atomizer Dry Scrubber SO_x/NO_x Process

A key feature in the new process is the existence of a so-called thermal window where simultaneous SO_x and NO_x removal can be obtained. The process is operated with an outlet temperature of the spray dryer absorber of 195 - 215°F. As in the normal dry scrubbing process lime is the main reagent. The process is operated in a recycle mode, i.e. recirculated dry waste product and slaked lime are mixed in the feed preparation system to a slurry with a solids content of 30-40%, before being atomized in the spray dryer absorber module. An inexpensive additive is required to make the process work in the mentioned thermal window. An additive, such as sodium hydroxide, is simply added into the feed preparation system in a quantity of approximately 10% of the lime added. In the Niro Atomizer spray dryer absorber (SDA) the efficient atomization and intimate mixing with the flue gas create a uniform recycle product which under close control is impregnated with a minimal amount of sodium sulfite formed by the selective reaction of sodium hydroxide with SO_2. Hereby a effective solid sorbent is created for the simultaneous removal of SO_x and NO_x. A major portion of the SO_x is removed in the SDA, but only minor amounts of NO_x is absorbed. The conditioned flue gas and activated particulates then enter the fabric filter where simultaneous removal of NO_x and SO_x is accomplished.

RESULTS OF BENCH SCALE AND PILOT PLANT INVESTIGATIONS

BENCH SCALE INVESTIGATIONS

As NO has very little affinity to water the SO_x/NO_x research program was directed towards developing active solid materials that would promote an SO_x/NO_x gas/solids reaction.

Bench scale tests were performed using a fixed bed reactor with a layer of reactive powder through which the flue gas was passed (Figure 2). The powder used was the normal dry scrubber waste solids including calcium sulfite, calcium sulfate and calcium hydroxide. The thickness of the powder layer and the gas velocity through the bed was adjusted to simulate as closely as possible the conditions in a baghouse filtercake.

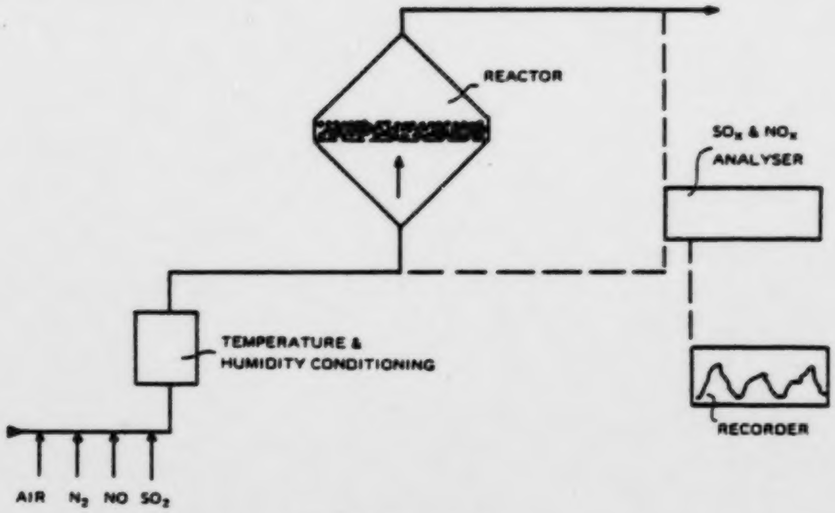

Figure 2. Bench Scale Set Up For SO_x/NO_x Tests

Effect of Additive

Various additives including sodium sulfite, Fe^{++}-compounds and Ethylenediaminetetra-acetic acid, disodium salt (EDTA) were used to impregnate the bed of powder. Sodium sulfite was found, for technical and economical reasons, to be the preferred reagent.

Tests with a soda ash based dry scrubber waste, which has the sodium sulfite in situ, showed no substantial NO_x absorption.

Effect of Reaction Temperature

Normal lime based dry scrubber powder was impregnated with small amounts of sodium sulfite and was subsequently exposed in the fixed bed reactor to various dry scrubber operating temperatures. The results shown in Figure 3 clearly indicate the existence of the so-called thermal window where SO_x and NO_x react simultaneously.

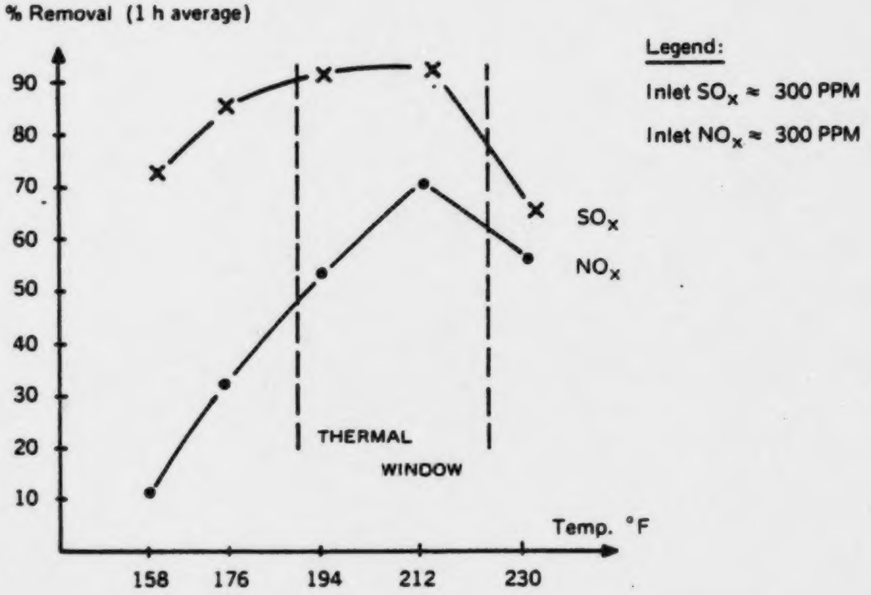

Figure 3. Influence of Reaction Temperature on SO_x/NO_x Reaction

Effect of SO_x/NO_x Ratio

The process is a simultaneous removal of SO_x and NO_x. If no SO_x is present no NO_x removal is achieved. The ratio between SO_x and NO_x should preferably be 1 (one) or greater. Table 1 shows the influence of SO_x/NO_x ratio.

TABLE 1. INFLUENCE OF SO_x/NO_x RATIO ON NO_x REMOVAL

SO_x PPM	NO_x PPM	SO_x/NO_x RATIO	NO_x REMOVAL (1st hour)	SO_x REMOVAL (1st hour)
0	300	0	0	-
100	300	1/3	34	92
300	300	1	70	91

Effect of Other Parameters

Other parameters such as the flue gas oxygen content and the moisture content of the powder also affect the SO_x/NO_x reaction. Table 2 & 3 show the results. Under normal power plant operating conditions the flue gas will contain 5% O_2 or more, so no negative effect on the SO_x/NO_x reaction is expected.

TABLE 2. INFLUENCE OF FLUE GAS OXYGEN ON SO_x/NO_x REACTION

O_2 % FLUE GAS	NO_x (1st hour)	SO_x (1st hour)
0.15	0	80
0.80	47	92
1.5	56	92
5.0	70	91

Table 3 shows that a certain residual moisture content has to be present in the powder to achive a substantial SO_x/NO_x removal.

TABLE 3. INFLUENCE OF POWDER MOISTURE ON SO_xNO_x REACTION

MOISTURE	NO_x	SO_x
MOIST POWDER	70	91
DRY POWDER	30	34

PILOT PLANT TEST RESULTS

Following the bench scale investigation a comprehensive pilot plant test program was carried out at the Niro Atomizer in-house FGD facility in Copenhagen. This pilot plant is routinely used to develop dry FGD system design parameters and provides very accurate scale-up information.

Table 4 (10) shows the correlation between pilot plant and full scale test results.

TABLE 4.

COMPARISON OF PILOT AND FULL SCALE (RIVERSIDE) TEST RESULTS***

PLANT	TEST No.	OPERATIONAL PARAMETERS									RESULTS	
		TEMPERATURES (°F)										
		PPM SO_2	SDA IN	SDA* OUT	BH** OUT	T Adsat.	GRT Sec.	FEED SOLIDS (%)	AIR TO CLOTH RATIO	CLEAN-ING CYCLE	% SO_2 REMOVAL TOTAL	SR
PILOT	6110-6115	700	314	144	136	18.5	12.9	33.5	1.25	1 hr	96.3	1.02
RIVERSIDE	20189	710	315	143	142	18.7	13.2	31.1	1.15	1 hr	96.8	1.09
PILOT	6120-6123	768	343	146	140	18.2	10.2	32.0	1.58	1 hr	97.7	1.15
RIVERSIDE	20186	758	340	146	146	18.7	10.5	32.6	1.44	1 hr	96.4	1.20

* SDA: Spray Dryer Absorber ** BH: Baghouse

*** Pilot Plant: 5,000 ACFM - Riverside Full-Scale Plant: 500,000 ACFM

The existence of the thermal window was confirmed during the pilot tests. Table 5 shows results from parametric tests that were not performed under optimum conditions with respect to other parameters.

TABLE 5.

PILOT PLANT DEMONSTRATION OF THERMAL WINDOW IN SO_x/NO_x PROCESS

SPRAY DRYER ABSORBER				% NO_x REMOVAL		% SO_x REMOVAL	
INLET			OUTLET				
TEMP. °F	SO_x PPM	NO_x PPM	TEMP. °F	SDA	SDA+BH	SDA	SDA+BH
330	1500	500	151	0	0	72	100
			194	6	21	54	83
			214	8	35	35	67
			230	5	37	25	55
			266	5	30	15	34

These tests with the spray dryer absorber / baghouse combination also reveal that the major part of the SO_x/NO_x reaction takes place in the filter cake on the baghouse bags. Only minor amounts of NO_x are removed in the spray dryer absorber.

Pilot plant results confirmed the influence of the ratio between SO_x and NO_x removal. Table 6 shows typical results.

TABLE 6. EFFECT OF SO_x/NO_x RATIO

SO_2 LEVEL INLET SDA PPM	1000	1600	2500
NO_x LEVEL INLET SDA PPM	300	300	300
% NO_x REMOVAL	40	50	65

As can be seen from the table the highest NO_x removal efficiency is - at the present state of development - achieved with higher sulfur coals. Pilot plant test results for typical high sulfur conditions are shown in Table 7. These tests were performed with addition of sodium hydroxide to the feed tank. A quantity of 10 weight percent in relation to the lime flow was used.

TABLE 7.

PILOT PLANT RESULTS WITH SO_x/NO_x PROCESS APPLIED TO HIGH SULFUR COAL

		TEST No. 12138	TEST No. 12149	TEST No. 12150
% Solids in Feed Slurry		34	30	36
Inlet SDA	Temp. °F	371	336	334
	SO_x PPM dry	2814	2762	2638
	NO_x PPM dry	310	309	346
Inlet BH	Temp. °F	188	222	204
% NO_x Removal		65.5	62.5	57.2
% SO_x Removal		98.8	86.2	93.3
Lime Stoichiometry		1.44	1.10	1.20

REACTION MECHANISM OF NEW SO_X/NO_X PROCESS

Based on the absorption results and chemical analyses of the SO_X/NO_X dry scrubber product, the reaction mechanism is illustrated in Figure 4.

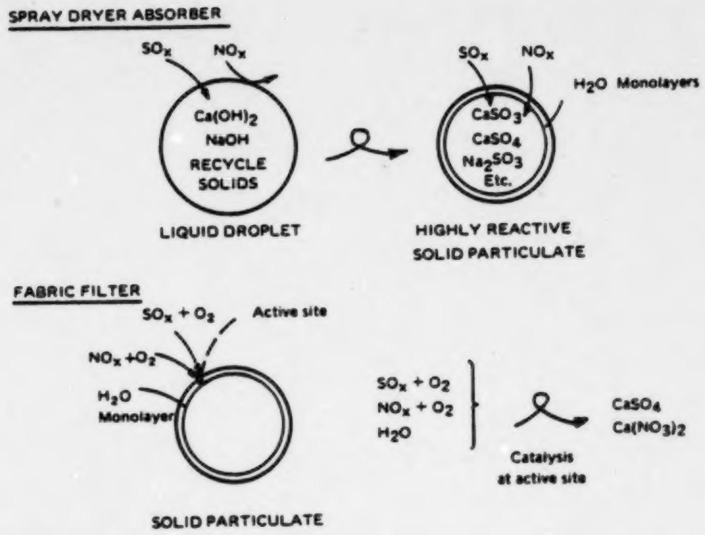

Figure 4. SO_X/NO_X REACTION MECHANISM

The added sodium hydroxide is converted to water soluble sodium sulfite during the absorption process in the spray dryer absorber. The sodium sulfite is partly precipitated as amorphous solids and partly kept in solution in the residual moisture, thus impregnating the spray dried solid particle of recycle solids and calcium hydroxide reaction products. Figure 5 shows a scanning electron microscope (SEM) photograph of such a particle.

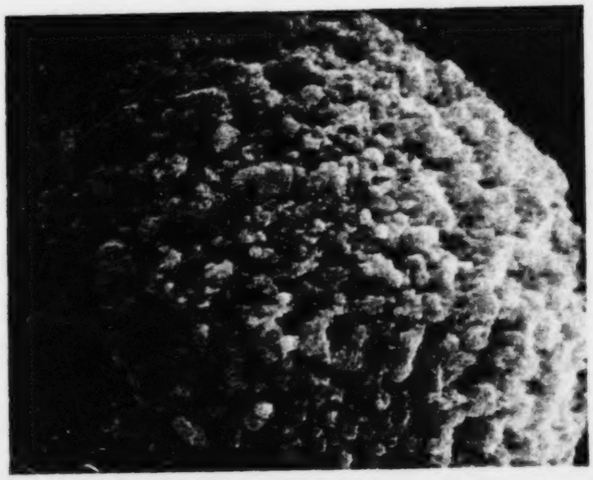

Figure 5. SEM-PHOTO OF ACTIVE SITES

The solid particulate with the active sites enters the fabric filter and forms a highly porous filter cake. According to Klingspor (11) between one and two monolayers of water will still exist on the solids surface in the thermal window where the process is operated. No NO_x absorption could be accomplished if too many monolayers were present. If on the other hand no water was present virtually no SO_2 absorption could be achieved.

At the active site SO_x and NO_x are oxidized producing calcium sulfate, calcium nitrate and calcium hydrogen sulfate, nitrate.

The proposed reaction mechanism is based on the following observations:

- Oxygen and water are necessary to obtain NO_x removal.
- The amount of NO_x absorbed agrees with the amount of nitrate found in the reaction product.
- The amount of sulfite in the reaction product is increased when compared with normal SO_x dry scrubber products.

RESULTS OF FULL-SCALE DEMONSTRATIONS

ARGONNE DEMONSTRATION

Since November 1981, i.e. for 2 years, Argonne National Laboratory has operated the first commercial JOY/NIRO dry scrubber designed for high (3.5%) sulfur coal (12, 13). The scrubber treats flue gas from a Spreader-Stoker boiler rated at 170.000 lb/hr of steam. The coal is from the Illinois basin. A typical analysis is shown in Table 8. System characteristics with respect to normal operation of the spray dryer absorber and the baghouse are given in Table 9.

TABLE 8

COAL CHARACTERISTICS FOR ARGONNE HIGH SULFUR COAL

COAL PARAMETER		VALUE
Heating value	Btu/lb	12,027
Moisture	%	9.59
Ash	%	7.40
Carbon	%	66.98
Hydrogen	%	4.65
Nitrogen	%	1.48
Sulfur	%	3.32
Chlorine	%	0.06
Oxygen	%	6.52

TABLE 9

ARGONNE DRY SCRUBBER SYSTEM CHARACTERISTICS

SPRAY DRYER ABSORBER

- Niro Atomizer 24.5 ft diameter spray dryer with compound gas disperser
- 35% feed slurry
- 20°F approach to saturation in exit gas
- 80 % SO_2 removal with lime stoichiometries of 1.0

BAGHOUSE

- JOY four compartment pulse flow baghouse
- 12 ft long woven fiber glass bags
- Air/cloth ratio 3:1
- Δ P: 3 inch W.G.

Argonne National Laboratories permitted Niro Atomizer to use their dry scrubbing system for a two week demonstration of the new SO_x/NO_x process. The demonstration was accomplished in April 1983 just before the boiler was brought down for its annual overhaul.

For the demonstration period Niro Atomizer brought in a continuous NO_x-monitor consecutively taking samples from spray dryer inlet, spray dryer outlet and baghouse outlet. During the tests back-up wet sampling of NO_x and SO_x were was Additive sodium hydroxide solution was stored in an intermediate 4000 gallon tank. A small centrifugal pump, a flow meter and some piping were all what were required to provide the capability of a controlled addition of sodium hydroxide to the recycle mixing tank.

Some mechanical problems - caused by the handling characteristics of the powder during the test period - showed the minor changes that have to be done on full-scale installations operating in SO_x/NO_x mode.

The test period was commenced with a test under normal operating conditions for the dry scrubber. No sodium hydroxide was added. This test demonstrated that under normal dry scrubber operating conditions virtually no NO_x is removed in the system.

Then the system was brought into the SO_x/NO_x scrubbing mode by increasing the spray dryer outlet temperature from 140°F to 195°F and starting sodium hydroxide addition to the recycle mixing tank. The result was very impressive and encouraging in as much as substantial NO_x removal was observed. Table 10 shows the overall performance with and without NO_x removal. In the SO_x/NO_x operating mode more than 50% NO_x removal was demonstrated while the scrubber was still removing 90%+ SO_2. The demonstration shows full agreement between pilot plant data and full-scale scrubber performance.

TABLE 10

DRY SCRUBBER SO_x/NO_x PERFORMANCE IN FULL-SCALE

MODE OF OPERATION	SCRUBBER					BAGFILTER OUTLET		RESULTS		
	INLET				OUTLET			% REMOVAL		STOICHIOMETRY FOR LIME
	PPM SO_2	PPM NO_x	TEMP. °F	GAS-FLOW lb/h	TEMP. °F	PPM SO_2	PPM NO_x	SO_x	NO_x	
NORMAL SO_2 REMOVAL ONLY	1800	290	335	200,000	148	90	281	95	3	1.4
SO_x/NO_x MODE	1500	280	330	140,000	195	60	126	95+	55	1.3

RIVERSIDE DEMONSTRATION

A 3 week test period is planned at JOY/NIRO's Riverside demonstration plant this fall. These tests will furnish additional full-scale data for process optimization. Results will be published as soon as available.

CHARACTERIZATION OF SO_x/NO_x DRY SCRUBBER WASTE MATERIAL

The waste product from dry scrubbing systems has been thoroughly analyzed and characterized over the last 3 - 4 years (14, 15, 16, 17, 18 and 19). These studies have shown that the dry scrubber waste product is an excellent material for landfill due to its pozzolanic properties and low permeability. The leachate quality can meet EPA limits and therefore no liners would be required in properly designed landfills.

The waste product from the SO_x/NO_x process has the same appearance as other dry scrubber poducts, i.e. a fine-grained, dry free-flowing powder.

The chemical composition is of course somewhat different. Table 11 shows the chemical analyses of the waste product from the Argonne tests with and without NO_x removal. The Argonne product has low ash content as the boiler is Stoker fired and multi clones are installed up-stream of the scrubber.

TABLE 11

CHEMICAL COMPOSITION (wt%) OF WASTE PRODUCT WITHOUT AND WITH NO_x REMOVAL

	SO_x REMOVAL	SO_x/NO_x REMOVAL
Sulfite as $CaSO_3 \cdot \frac{1}{2}H_2O$	59	16
Sulfate as $CaSO_4 \cdot 2H_2O$	16	43
Nitrate as $Ca(NO_3)_2$	~0	3
Excess alkalinity as $Ca(OH)_2$	15	21
Ash & Inerts	9	8
Moisture	1	1
Na_2SO_3	~0	9

As can be seen from the analyses, a major part of the sulfite has been oxidized to sulfate when the scrubber operates in SO_x/NO_x mode. The absorbed NO_x is present in the product as nitrate. No nitrite has been found.

To evaluate the suitability of the waste product for landfill the following important parameters were investigated:

- Density
- Compressive strength
- Permeability
- Leachate quality

Figure 6 shows the moisture-density curve for the two products. The SO_x/NO_x product has a higher dry density and about the same optimum moisture content as the normal dry FGD product.

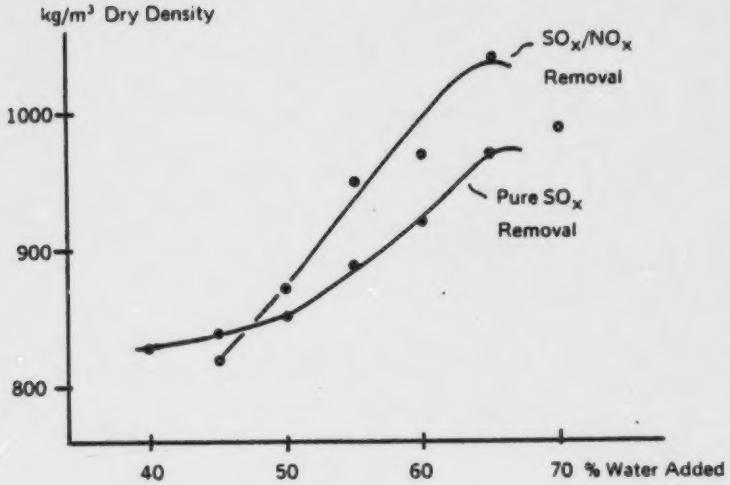

Figure 6. Moisture-density data of waste product from pure SO_x and combined SO_x-NO_x-removal

Table 12 shows a comparison of compressive strengths and permeabilities. The strength development is the same as for the normal product, and the permeability is very low for both products.

TABLE 12

COMPRESSIVE STRENGTH AND PERMEABILITY OF PURE SO_x AND COMBINED SO_x/NO_x PRODUCTS *

COMPRESSIVE STRENGTH (kPa)	PURE SO_2 PRODUCT	SO_x/NO_x PRODUCT
1 DAY	120	110
7 DAYS	180	200
30 DAYS	480	340
56 DAYS	600	580
PERMEABILITY (10^{-6} cm/s)		
1-2 DAYS	2.5	6
28 - 30 DAYS	-	-

* All samples are compacted to 960-980 kg/m^3 in dry density

The leachate quality obtained by the EPA Toxic Extraction Procedure is shown in Table 13. All of the leachate values shown are less than 100 times the Primary Drinking Water Standards (PDWS), therefore the product would most likely be classified as non-hazardous.

TABLE 13

LEACHATE QUALITY OF WASTE PRODUCTS FROM PURE SO_x AND COMBINED SO_x/NO_x REMOVAL*

	EPA-PDWR mg/l	PURE SO_2 REMOVAL mg/l	SO_x/NO_x REMOVAL mg/l
ARSENIC	0.05	0.001	0.001
BARIUM	1.0	< 0.5	< 0.5
CADMIUM	0.010	< 0.005	< 0.005
CHROMIUM	0.05	0.010	0.012
LEAD	0.05	< 0.05	< 0.05
MERCURY	0.002	< 0.002	< 0.002
NITRATE (as N)	10	2	260
SELENIUM	0.01	0.004	0.004
SILVER	0.05	< 0.01	< 0.01

* test carried out according to EPA toxic extraction procedure

The waste product from the Argonne tests contains little ash. The dry scrubber product from a typical high sulfur application would have a fly ash content of 50 - 60 %. This would change the characteristics in a positive direction.

It is expected that the SO_x/NO_x waste product will be regulated as non-hazardous. However, due to somewhat elevated values for nitrate and total dissolved solids (TDS), it may be necessary to install a liner before disposal.

Currently Niro Atomizer is working on methods to improve leachate quality and to reduce leachate quantity.

COMPARISON OF ECONOMICS WITH STATE OF THE ART SO_x/NO_x PROCESSES

A cost comparison has been made between the Niro Atomizer dry SO_x/NO_x process and a conventional limestone wet scrubber / selective catalytic de-NO_x (SCR) system. The reason for using SCR for comparison and not the cheaper selective non-catalytic de-NO_x system (SNCR) is that the latter is able to achieve only 30 - 50% NO_x removal. Furthermore SNCR has not been demonstrated on high sulfur coal. Excessive ammonia carryover is likely to result in severe air preheater plugging problems as well as problems with waste characteristics. This excludes the use of SNCR for high sulfur applications (22, 23).

The basis for the cost comparison is a 500 MW power plant burning 3% sulfur coal and emitting 300 ppm NO_x. A 65% NO_x removal and a 90% SO_2 removal are assumed for both systems. The operating cost for the SCR system is critically dependent on the catalyst lifetime. A two year lifetime has been assumed for this cost comparison.

The capital requirement and the operating/maintenance cost for the two systems are shown in Table 14. It is seen that the capital requirement is considerably less for the Niro Atomizer dry SO_x/NO_x system. Operating and maintenance costs are very similar for the two systems.

TABLE 14

OPERATING AND CAPITAL COST ESTIMATE FOR COMBINED SO_x AND NO_x REMOVAL SYSTEMS *

COST ITEM	WET LIMESTONE + SCR	NIRO DRY SO_x/NO_x
CAPITAL REQUIREMENT	105 M$ FGD 30 M$ SCR 135 M$	85 M$
OPERATING AND MAINTENANCE COSTS (1st Year) (O&M)	17.2 M$ FGD 4.6 M$ SCR 21.8 M$	19.5 M$
LEVELIZED CAPITAL REQUIREMENT	7.4 mills/kWh	4.8 mills/kWh
LEVELIZED O&M	12.7 mills/kWh	11.3 mills/kWh
TOTAL LEVELIZED COST	20.1 mills/kWh	16.1 mills/kWh

* References: (3, 20, 21 and 22)

A single annual levelized cost value has been computed based on a 30 year economic plant life to represent the revenue requirement for the operating and maintenance (O&M) cost and the capital requirement . For the O&M cost a levelization factor of 1.9 is used. For the capital requirement a levelization factor of 0.18 is used. The total levelized cost for the Niro Atomizer dry SO_x/NO_x system is 16.1 mills/kWh compared to 20.1 mills/kWh for the limestone wet scrubber/SCR system.

Besides having favourable economics when compared to a limestone wet scrubber/SCR system the Niro Atomizer dry SO_x/NO_x process has the additional advantage of being one integral air pollution control system. It will be easier to operate and will not interfere with the boiler operation since no equipment is installed upstream the air preheaters.

CONCLUSION

- Dry scrubbing has gained much acceptance within the utility industry due to its low cost and ease of operation.
- Dry scrubbing with simultaneous removal of SO_x and NO_x is a new process - developed by Niro Atomizer - which uses the same equipment as provided for dry scrubbers presently in operation or under construction.
- The process is well suited for high sulfur coal applications where low NO_x burners have reduced NO_x to the 0.25 - 0.35 lb/MBtu level. The process can further reduce NO_x emissions down to the 0.1 lb/MBtu level together with 90 - 95% SO_x removal.
- Process performance has been tested in pilot-scale and has been confirmed in full-scale at Argonne National Laboratories.
- Waste products from the process will most likely be classified as non-hazardous.
- Economics of the new process compare favourably with limestone wet scrubbing plus selective catalytic reduction (SCR).

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PROCESS CHARACTERIZATION
OF SO_2 REMOVAL IN SPRAY ABSORBER/BAGHOUSE SYSTEMS

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ABSTRACT

A new equation for correlating the SO_2 removal with stoichiometric ratio, approach to saturation, inlet SO_2 concentration, inlet temperature, and inlet moisture content is proposed. It reveals essential information about the kinetics of the SO_2 removal process. The chemical reaction between the SO_2 and lime appears to limit the rate of SO_2 removal in the spray absorber. The effect of recycle enhancement of SO_2 removal may be understood in terms of this model. The model also permits a classification of flyash suitable for designing dry flue gas desulfurization (DFGD) systems.

DFGD pilot plant results are reviewed from theoretical and mechanistic viewpoints. The SO_2 removal as a function of the gas flow rate (or average residence time) and the angle of the secondary swirl vanes in the absorber gas disperser is discussed relative to the velocity flow field in the absorber. The SO_2 removal efficiency is also discussed as a function of the atomizer angular speed.

When the gas flow rate and approach to saturation are fixed, the SO_2 removal efficiency in the spray absorber increases with increasing stoichiometric ratio and inlet temperature and decreasing inlet moisture content and SO_2 concentration. An explanation of these effects is given by their impact on the droplet diameter integrated over the lifetime of the droplets. Another effect that increases the time-integrated droplet diameter is the decrease in the volume of the core of solids (lime) for smaller slurry droplets.

Partial recycle of the spent products, in the absence of externally injected flyash, results in SO_2 removal efficiency being nearly independent of SO_2 concentration. Thus, the enhancement from recycle is greater for the higher SO_2 concentrations. SO_2 removal performance with recycle was improved by the injection of flyash derived from a low sulfur Texas lignite and from high sulfur eastern coals. The enhancement with recycle in the latter case is dramatic and supports DFGD applicability to high sulfur coal.

NOMENCLATURE

ε_s = SO_2 removal efficiency across spray absorber

α = $\gamma_s C_p/(N_{Nu}K)$

γ_s = Product deposition reaction rate constant

C_p = Flue gas specific heat at constant pressure

N_{Nu} = Nusselt number for liquid droplets in gas stream

K = Thermal conductivity of flue gas

X_c = Correction factor for effect of the solids fraction on the SO_2 removal rate

$= 1 - D_f^3 / D_i^3$

D_i = Initial mean droplet diameter

D_f = Final mean particle diameter of evaporated droplet

S = Stoichiometric ratio

T_{si} = Spray absorber inlet temperature

T_{sf} = Spray absorber outlet temperature

T_{as} = Adiabatic saturation temperature of flue gas entering spray absorber

A = Time-integrated droplet diameter

$= \Gamma \int_0^{\tau} D(t)dt$

Γ = Concentration of droplets

τ = Droplet evaporation time

$D(t)$ = Droplet diameter at time, t

u = Empirical constant characteristic of a given flyash

F = Mass fraction of solids in slurry droplet or atomizer slurry flow

P = Reduced penetration

$$= - \ell n(1 - \varepsilon_s) / \{X_c \; S \; [\ell n \frac{T_{si} - T_{as}}{T_{sf} - T_{as}}]\}$$

INTRODUCTION

The New Source Performance Standards (NSPS) for SO_2 require cleanup of flue gases from utility boilers burning even low-sulfur coals. The parallel increase in application of fabric filters to control particulate emissions from coal-fired boilers has generated considerable technical and commercial interest in dry flue gas desulfurization (DFGD) processes. The most commercially promising DFGD process is spray absorption. In this process, an alkali sorbent liquid is atomized in a spray dryer using flue gas as the drying medium. The DFGD reactions occur between the gas and droplets in intimate contact during the various stages of drying.

Eventually the reaction products are reduced to a suspended solid particulate in the gas stream, some solids dropping out in the absorber hopper. When a fabric filter is used for removal of remaining spent sorbent and flyash from the gas stream, further DFGD reactions occur in the filter cake by virtue of unreacted alkali and residual moisture. While a variety of calcium- and sodium-based alkalis have been evaluated, lime currently has the greatest commercial importance.

In 1978, the Buell Emission Control Division of the Envirotech Corporation (now part of General Electric Environmental Services, Inc.) and Anhydro A/S constructed a spray dryer FGD prototype facility at the Martin Drake Station of the City of Colorado Springs. A 4.01-m^3/s (8,500-cfm) spray absorber was installed in slipstream configuration in the inlet duct to a 189 m^3/s (400,000 cfm) fabric filter serving an 85 MW pulverized-coal-fired steam generator burning northwestern Colorado coal. An extensive program of parametric performance and process demonstration tests was undertaken at that facility jointly with EPA and the City of Colorado Springs. As a result improved models for spray dryer FGD performance were developed. This paper presents these models and experimental data which support their validity.

THEORY

Two models for predicting the SO_2 removal in the spray absorber/baghouse system are presented in this paper. The first is an analytical model restricted to straight-through operating conditions only. The second is an empirical model applicable under wet recycle of solids collected from the baghouse and/or the spray absorber hoppers. The analytical model provides a theoretical basis for understanding the fundamental processes which remove SO_2 by spray absorption. The empirical model extends the usefulness of the analytical model by including the effect of flyash on SO_2 removal by spray absorption.

Analytical Model for Straight-Through SO_2 Removal (No Recycle)

An analytical expression for the SO_2 removal rate (or efficiency) in a spray absorber may be derived for the following idealized conditions within the spray absorber.

1) The gas velocity is uniform within the spray absorber (the presence of vortices, eddies, and recirculation zones is ignored).

2) The droplet/gas mixing is uniform, with the droplets following gas streamlines.

3) Most SO_2 removal takes place during the period of water evaporation from the droplets (often called Phase I), when surfaces are maintained at the adiabatic saturation temperature.

4) The rate of SO_2 removal is determined by the rate of the chemical reaction between the SO_2 and lime to form sulfites or sulfates.

Under these assumptions, the simultaneous occurrence of SO_2 removal and water evaporation is controlled by the rate of SO_2 removal, the rate of droplet size changes, and the conservation of thermal energy in the spray absorber (1,2). Consistent with these relationships the SO_2 removal efficiency, under straight-through conditions, can be expressed by Equation 1.

$$- \ell n(1 - \varepsilon_s) = \alpha \, X_c \, S \, [\ell n \frac{T_{si} - T_{as}}{T_{sf} - T_{as}}] \qquad (1)$$

The correction factor, X_c, arises from the proportionality between the penetration [i.e., $- \ell n(1 - \varepsilon_s)$] and the time-integrated droplet diameter. For pure water droplets, the final diameter, D_f, = 0. When solids are present in the slurry droplet, its initial diameter is increased due to volume displacement by the solids, and its final diameter, at the end of evaporation, is non-zero due to the volume of the solids. As the solids fraction increases, the departure of the correction factor, X_c, from unity becomes progressively more significant. The correction factor, X_c, depends on both the mean diameter, D_i, of droplets produced in the atomizer and the final mean diameter, D_f, of particles produced at the end of evaporation. Its accurate evaluation is complicated by the dependence of the initial droplet diameter on the solids fraction in the slurry being atomized and by the dependence of the final particle diameter on the extent to which the dried particle retains some moisture. A reasonable expectation is for $X_c \simeq 1$ for solids fractions <0.1. At solids fraction of 0.5, the correction factor, X_c, appears to be about 0.5 to 0.7. Table 1 discusses the dependence of the model prediction for the spray absorber SO_2 removal efficiency on selected spray absorber and flue gas parameters.

Empirical Model for SO_2 Removal with Wet Recycle

The SO_2 removal efficiency in the presence of wet recycle of offproduct is invariably better than that under straight-through conditions. In addition to unspent lime, the recycled offproduct contains flyash. The flyash appears to have a synergistic effect on the utilization of lime from both fresh and recycled sources. Different types of flyash in the recycle slurry provide different levels of enhancement of the SO_2 removal. This recycle enhancement appears to be related to the coal from which the flyash originates and on flyash characteristics such as pH of hydrolysis, particle size, and chemical composition. Since the flyash effect on SO_2 removal through wet recycle is not well understood, a fundamental relationship for the efficiency in the presence of recycle cannot be derived. The following empirical relationship is suggested, based on Equation 1, to be consistent with the observation of improved SO_2 removal in the presence of recycle which increases the solids fraction, F , in a newly produced slurry droplet (or in the atomizer feed):

TABLE 1. MODEL PREDICTIONS FOR OPERATIONAL EFFECTS ON STRAIGHT-THROUGH REMOVAL IN SPRAY ABSORBER

Parameter	How Parameter Enters Equation 1	Relationship to SO_2 Removal	Comment
Absorber Residence Time	Does not enter equation explicitly.	None, provided residence time is long enough to allow complete evaporation.	Water flow rate and the inlet and outlet gas temperatures are held at conditions which give complete evaporation and maintain $X_c = 1$.
Atomizer Angular Speed	Does not enter equation explicitly.	None, provided droplets are of such size that they evaporate completely.	For a fixed atomizer water flow rate, the time-integrated droplet diameter is independent of the initial droplet diameter, provided complete evaporation occurs and $X_c \simeq 1$.
Gas Inlet Temperature	Enters equation explicitly as T_{gi}.	SO_2 removal efficiency increases with increasing inlet temperature for fixed approach to saturation temperature.	The water required to reach the same approach to saturation temperature increases with increasing inlet gas temperatures. The time-integrated droplet diameter increases with increasing inlet gas temperature.
Gas Inlet Moisture Content	Enters equation indirectly through the adiabatic saturation temperature, T_{as}.	SO_2 removal efficiency decreases with increasing inlet moisture content for fixed inlet gas temperature and approach to saturation.	For the same inlet gas temperature, the adiabatic saturation temperature increases with increasing inlet moisture content. Consequently, the water required to reach the same approach to saturation at fixed inlet gas temperature decreases with increasing inlet moisture content.
Inlet SO_2 Concentration	Enters equation indirectly through the correction factor, X_c.	SO_2 removal efficiency is independent of inlet SO_2 concentration for low concentrations (<1000 ppm). SO_2 removal efficiency decreases with inlet SO_2 concentration at high concentrations (>1000 ppm) at fixed stoichiometry.	Water flow rate and the inlet and outlet gas temperatures are held at conditions which give complete evaporation. The droplet evaporation time decreases with increasing inlet SO_2 concentration. For inlet SO_2 concentrations less than about 1000 ppm, the lime rates are low such that $X_c \simeq 1$. At higher inlet SO_2 concentrations, X_c decreases with increasing concentration.

$$- \ell n(1 - \varepsilon_s) = X_c\ S\ [\ell n \frac{T_{si} - T_{as}}{T_{sf} - T_{as}}]\ (u\ F + \alpha) \qquad (2)$$

or

$$P = u\ F + \alpha \qquad (3)$$

where u is an empirical constant which varies with flyash type. The constant, α, comes from Equation 1. Implications of this empirical equation will be discussed later.

EXPERIMENT

TEST OBJECTIVES AND EQUIPMENT

The overall objective of the dry FGD (DFGD) pilot program at Colorado Springs was to demonstrate and characterize SO_2 removal in a spray absorber/baghouse system for low sulfur and high sulfur applications. Although lime was the main reagent tested, other reagents, such as trona and dolomitic lime, which may be preferable for certain power plant sites, were also evaluated. Limestone was also tested because its effective use in a DFGD system will dramatically shift FGD economic evaluation in its favor. Other secondary objectives of the Colorado Springs program were to study the effects on SO_2 removal of baghouse reheat, adipic acid addition to lime, and the use of wastewater for lime slaking. Details of the tests conducted in the pilot program are reported elsewhere (3,4,5,6). Only those tests which are relevant to the proposed models are discussed in this paper.

The primary elements of the Colorado Springs prototype plant are a 3.8-m (12.5-ft) diameter spray absorber equipped with a top-entry vaned scroll-type gas disperser; a centrifugal atomizer with a 0.25-m (10-in.) diameter wheel; a full-scale commercial-type lime storage, handling, and slaking system; and a reverse-air baghouse fitted with full-size (0.30-m or 12-in. diameter, 9.14-m or 30-ft length) fiberglass filter bags. The plant is also equipped with a four-element cyclone collector in parallel with the baghouse at the absorber outlet to provide the capability to independently vary the absorber flow volume and baghouse air/cloth ratio. An SO_2 vaporizing and injection system enabled increasing the SO_2 concentration at the spray dryer inlet from 200 to 400 ppm (the value typical of the Colorado coal) to values in the range from 1,000 to 4,000 ppm. The higher values in this range are characteristic of the flue gas derived from burning high sulfur eastern coals. Flue gas characteristic of a given coal other than the Colorado coal burned at the Martin Drake Station was simulated by the external injection of SO_2 as needed and flyash derived from a boiler firing the coal type of interest.

Major measurement systems included venturi meters for gas flow measurement at the absorber and baghouse inlets; an extractive SO_2 monitoring system (DuPont Model 460) with probes at the absorber inlet, at the absorber outlet, and at the baghouse outlet; a dew point hygrometer with a probe at the baghouse outlet; magnetic flow meters for the measurement of slurry flows; and extensive pressure and temperature instrumentation. All equipment and materials required for instrument calibration were available onsite, as were chemical laboratory facilities for measurement of slurry reactant concentration, slurry total solids by evaporation, and offproduct moisture content.

The Colorado Springs prototype plant was sized to accommodate a wide range of operating parameters. Following the plant start-up, shakedown, and initial optimization tests, operation was established within the range of parameters summarized in Table 2.

RESULTS

OPTIMIZATION TESTS

Average Gas Residence Time in Spray Absorber

Figure 1(a) summarizes the results of a series of tests to determine the variation in the SO_2 removal efficiency across the spray absorber, baghouse, and the spray absorber/baghouse system, due to variation in the residence time (from 7 to 12 seconds) for fixed atomizer angular speed (12,500 rpm), stoichiometric ratio (1.0), and approach to saturation (17°C, 30°F). The spray absorber SO_2 removal efficiency increases slowly with decreasing residence time (increasing flow rate). On the basis of the analytical model, a constant SO_2 removal efficiency with respect to residence time would be expected. The increase shown by test results appears to be due to improved gas flow distribution and gas/droplet mixing with increasing gas flow rate.

The SO_2 removal in the baghouse was more sensitive to the average gas residence time in the spray absorber than was SO_2 removal in the spray absorber itself. The baghouse removal shows a maximum at a residence time of about 10 seconds. The increase in the baghouse efficiency when the residence time increases from 7 to 10 seconds (decreasing gas flow rate) seems due to the decreased cyclonic action in the spray absorber and the consequent increase in the mass loading of the spent product into the baghouse. The decreasing baghouse performance with increasing spray absorber residence times beyond 10 seconds seems to be due to poorer gas flow velocity distribution and the accompanying reduced gas/particle mixing in the spray absorber. This baghouse behavior is also consistent with the improved cyclonic activity and the accompanying decrease in the mass loading to the baghouse when the angular momentum of the slurry becomes progressively greater at the low gas flow rates corresponding to residence times greater than 10 seconds.

Atomizer Wheel Angular Speed

Figure 1(b) summarizes the observed variation in the spray absorber, baghouse, and spray absorber/baghouse system SO_2 removal efficiencies with atomizer angular speed. The gas flow rate was fixed at 4.01 m^3/s (8500 cfm)

TABLE 2. COLORADO SPRINGS SPRAY ABSORPTION PROTOTYPE PLANT SUMMARY OF OPERATING PARAMETERS

Parameter[a]	Value			Maximum/Minimum	Variation during test, percent
	Typical	Minimum	Maximum		
ABSORBER					
Inlet flow volume, cfm[b]	8,500	5,000	10,600	2.1	5
Inlet temperature, °F	350	250	400	1.6	2
Inlet SO_2 concentration, ppm	1,000	250	4,000	16.0	2
Inlet saturation temperature, °F[c]	120	110	140	1.3	5
Outlet temperature, °F	140	122	180	1.5	3
T_p, approach temperature, °F[d]	20	12	40	3.3	5
BAGHOUSE					
Inlet flow volume, cfm	3,000	1,000	6,000	6.0	5
ATOMIZER					
Wheel angular speed, rpm	12,500	6,300	14,000	2.2	1
SLURRY SYSTEM					
Lime slurry flow, gpm[e]	1.0	0.5	3.0	6.0	2
Total atomizer feed, gpm	2.0	1.0	6.0	6.0	2
Lime slurry total solids, percent	15	8	50	6.3	2

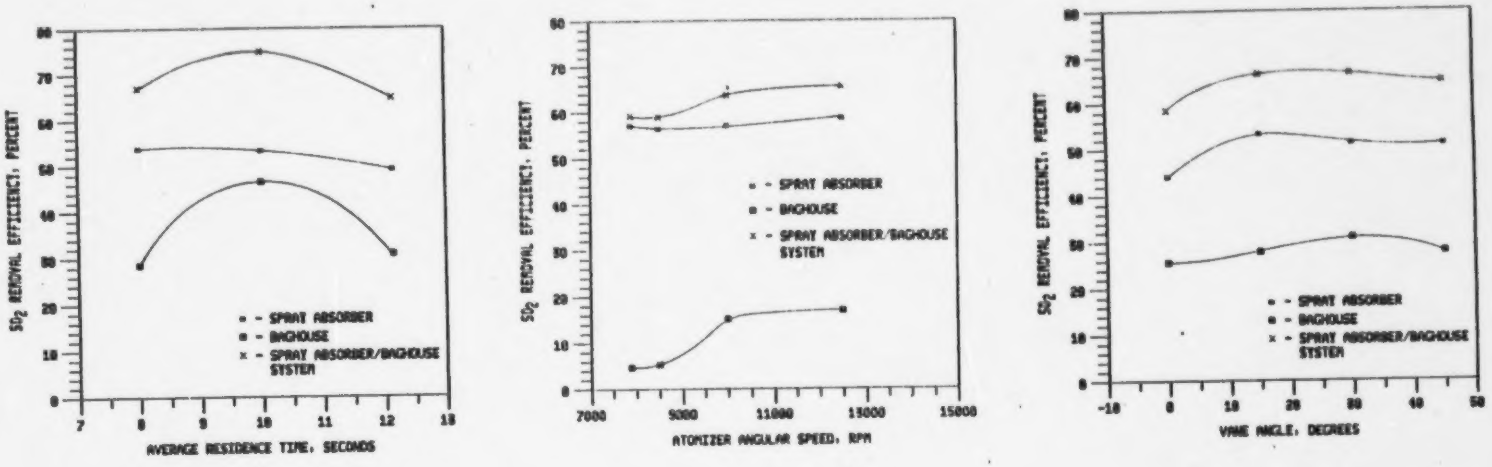
[a] Units shown apply only to columns under "Value."

[b] Multiply cfm value by 4.72×10^{-4} to convert to m^3/s.

[c] T, °C, = 5(T, °F − 32)/9.

[d] T_p, °C = 5(T_p, °F)/9.

[e] Multiply gpm value by 6.3×10^{-5} to convert to m^3/s.



1(a). Observed dependence of straight-through SO_2 removal efficiency on average gas residence time in the spray absorber. A residence time of 10 seconds appears to optimize the system SO_2 removal efficiency.

1(b). Observed dependence of straight-through SO_2 removal efficiency on the atomizer angular speed. The system SO_2 removal efficiency appears to be optimized for angular speeds $\geq$ 10,500 rpm.

1(c). Observed dependence of straight-through SO_2 removal efficiency on the inlet gas disperser vane angle (with the vertical). A vane angle of 15 to 20° appears to optimize the system SO_2 removal efficiency.

Figure 1. Optimization test results for pilot lime spray absorber/baghouse system without recycle of offproduct. [For these tests, the following parameters were fixed: inlet SO_2 concentration (1000 ppm), stoichiometric ratio (1.0), inlet gas temperature (177°C, 350°F), and approach to saturation (17°C, 30°F).]

(10-second average residence time), the stoichiometric ratio at 0.95, approach to saturation at 17°C (30°F), and the inlet temperature at 177°C (350°F) for these tests. SO_2 removal in the spray absorber was relatively independent of wheel speed in the range tested and, consequently, independent of droplet diameter (6,7,8). This independence is consistent with the theoretical expression summarized in Equation 1 for the case of low solids concentration in the atomizer feed. In this case, $D_f \simeq 0$ and the correction factor, $X_c, \simeq 1$. Hence the time-integrated diameter of the droplets to which the SO_2 removal is proportional is independent of initial droplet diameter.

Figure 1(b) also shows a stepwise increase in the baghouse SO_2 removal efficiency at an atomizer angular speed of about 10,500 rpm. The system SO_2 removal efficiency reflects the sensitivity of the baghouse SO_2 removal efficiency to variations in atomizer angular speed and appears to be optimized at atomizer angular speeds above 10,500 rpm.

Atomizer Disperser Vane Angle

By tilting the secondary swirl vanes of the gas disperser from the vertical (i.e., parallel to the atomizer wheel axis), it is possible to impart a tangential component to the gas velocity as the gas enters the spray chamber and thereby induce a swirl in the gas velocity flow field. An initial swirl in the gas velocity flow field is advantageous in extending the gas residence time by inducing a helical motion to the gas molecules in the spray absorber and in improving gas/particle mixing by providing a centripetal acceleration to the particles. Figure 1(c) summarizes the observed SO_2 removal efficiencies with lime as a function of the vane angle (measured from the vertical) for fixed stoichiometric ratio (0.92), inlet gas temperature (177°C, 350°F), approach to saturation (17°C, 30°F), and atomizer angular speed (12,500 rpm).

The vane angle appears to improve the SO_2 removal in the spray absorber noticeably in the range from 0 to 20°. Increasing the vane angle beyond 20° does not lead to noticeable improvement in the performance of the spray absorber. This suggests that a higher level of turbulence at the higher vane angle does not allow the swirl to be completely characterized by the initial angle of projection and that, as the flow progresses into the chamber, a spectrum of swirl angles is produced. In this way, the turbulence in the flow field removes any identity of the flow with the initial swirl angle for vane angles above 20° in the range from 0 to 45°. The contribution of the angular momentum of the slurry to the swirl of the gas will also tend to lessen the effect of increasing the vane angle.

The baghouse SO_2 removal efficiency appears to be independent of the vane angle. This suggests that the size distribution of particles produced at the end of evaporation of the slurry droplets created in the rotary atomizer are in such a range that their carryover to the baghouse is unaffected by changes in vane angle.

Spray Absorber Inlet Temperature at Fixed Inlet Flue Gas Moisture Content

The effect of varying the temperature of the flue gas at the spray absorber inlet without changing its moisture content was studied in a series of tests. In these optimization tests, the atomizer disperser vane angle was set at 15° and the atomizer wheel speed was maintained at 12,500 rpm. As in earlier tests, the gas flow rate was 4.01 m^3/s (8,500 cfm) and corresponded to an average residence time of 10 seconds in the spray absorber. The spray absorber inlet temperature was varied by using a surface heat exchanger with water as the cooling fluid upstream of the spray absorber inlet.

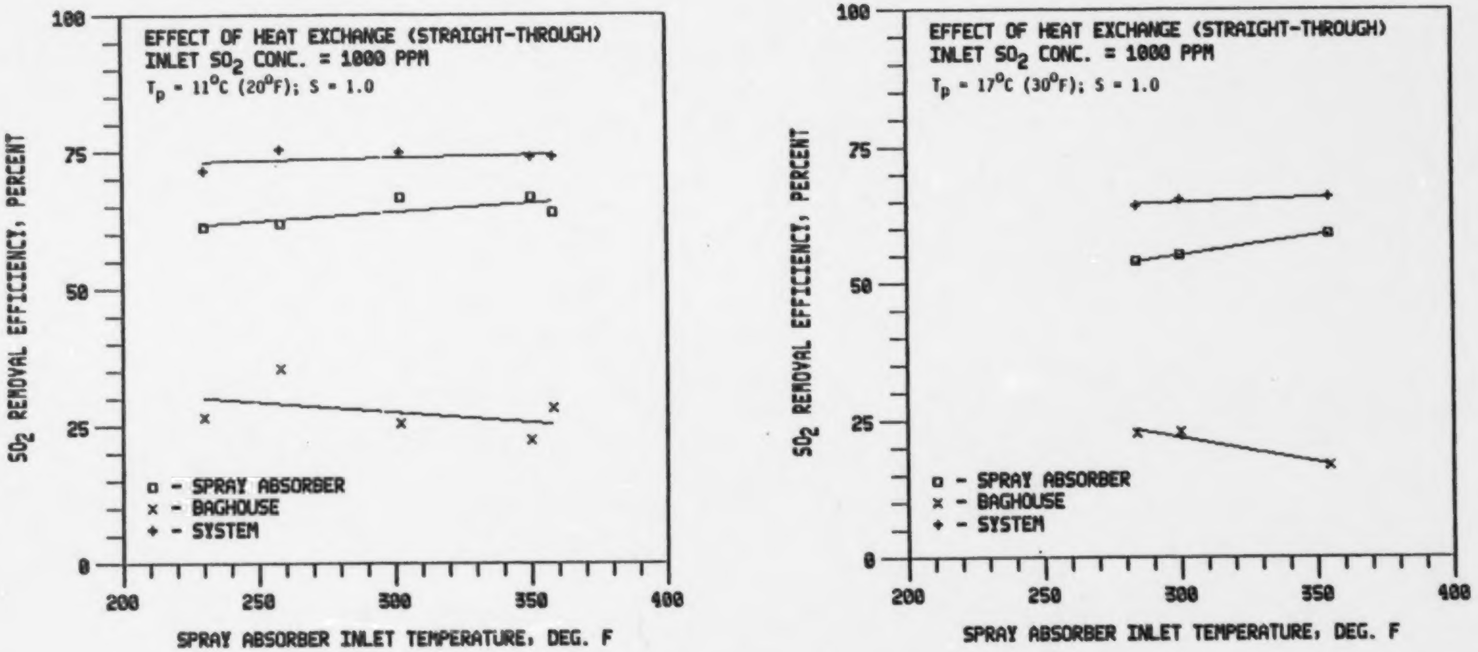
Figures 2(a) and 2(b) display the experimental results for the SO_2 removal efficiencies under straight-through conditions for approach temperatures of 11°C (20°F) and 17°C (30°F). The spray absorber SO_2 removal efficiency increased with increasing inlet temperature, while the baghouse efficiency displayed the opposite trend. Thus the system SO_2 removal efficiency increased only weakly with increasing inlet temperature. The rate of increase of the spray absorber efficiency and the rate of decrease of the baghouse efficiency with increasing inlet temperature at a 17°C (30°F) approach to saturation are more than double the rates at an 11°C (20°F) approach.

Since the thermal energy of the inlet gas decreases with decreasing temperature, the water required to reach a fixed approach to saturation decreases with decreasing inlet temperature. When the atomizer is operated at a fixed angular speed and the lime feed rate is low enough such that the correction factor, X_c, $\simeq 1$, the mean size of droplets produced in the atomizer may be regarded as being independent of the inlet gas temperature. The decrease of the atomizer water feed rate with decreasing inlet gas temperature then translates to a decrease in the initial total surface area of droplets produced in the atomizer because fewer droplets are formed.

The decrease of the spray absorber efficiency with a decrease in the inlet temperature can be related to this lower total initial surface area of droplets resulting from a lower water feed rate to the atomizer to reach the same approach to saturation. The corresponding increase in the baghouse removal suggests that the slower evaporation rate induced by the lower inlet temperature leaves particles at the spray absorber outlet with a higher moisture content, allowing more effective SO_2 removal in the baghouse. For the same gas flow rate at the same inlet temperature, the atomizer water requirement to reach the 17°C (30°F) approach is less than that for the 11°C (20°F) approach. Thus, the higher rate of decrease of spray absorber efficiency with decreasing inlet temperature for the 17°C (30°F) approach may arise from the difference in the water requirements and the associated difference in the initial droplet surface area.

Spray Absorber Inlet Temperature Variation with Prequench

The previous section investigated the effect on SO_2 removal of varying the inlet temperature while holding the absolute humidity fixed. That set of



2(a). Dependence of the straight-through SO_2 removal efficiency on the spray absorber inlet gas temperature varied by cooling the inlet gas with a surface heat exchanger for fixed approach to saturation (11°C, 20°F).

2(b). Dependence of the straight-through SO_2 removal efficiency on the spray absorber inlet gas temperature varied by cooling the inlet gas with a surface heat exchanger for fixed approach to saturation (17°C, 30°F).

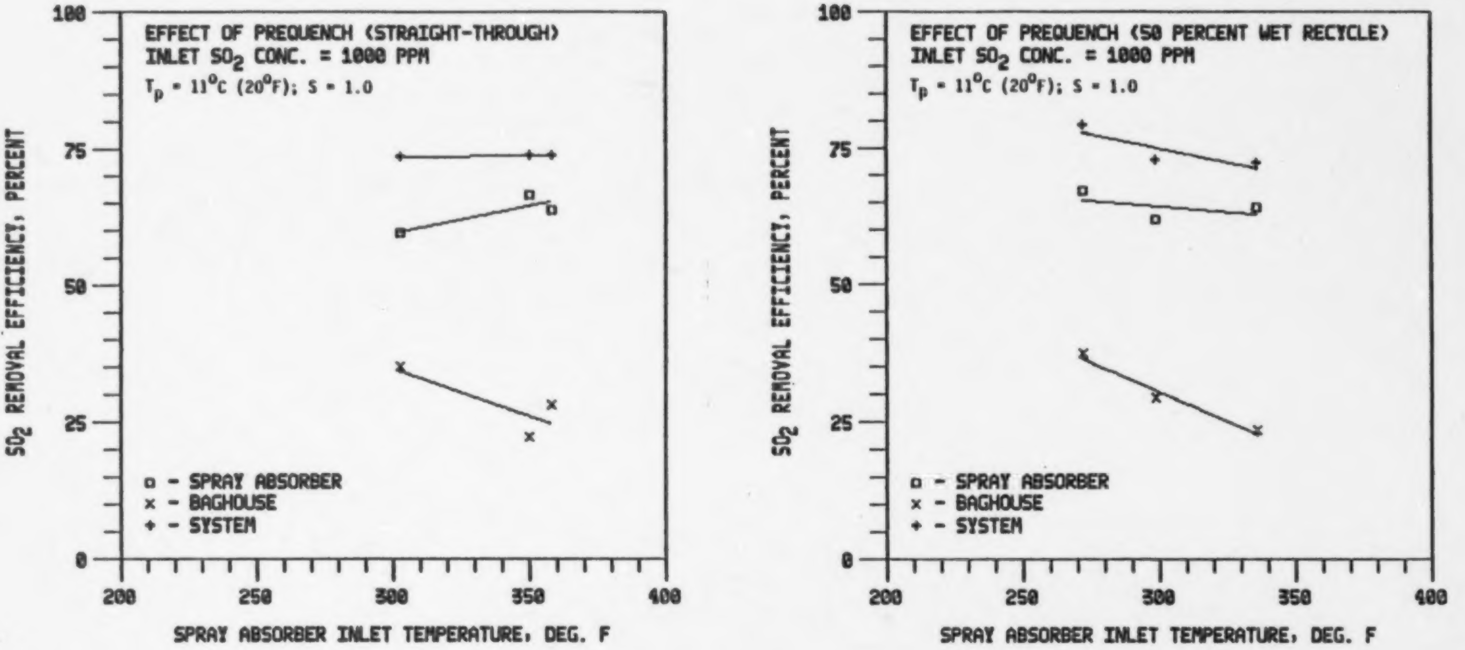
Figure 2. Effect of the spray absorber inlet gas temperature on the SO_2 removal efficiency in pilot lime spray absorber/baghouse system without recycle of offproduct. [For these tests, the following parameters were fixed: inlet SO_2 concentration (1000 ppm), stoichiometric ratio (1.0), average gas residence time in spray absorber (10 seconds), atomizer angular speed (12,500 rpm), and inlet gas disperser configuration (20° vane angle with vertical and small vane inserts).]

tests was useful in understanding the effect of inlet temperature on spray absorber/baghouse performance for bituminous and subbituminous coals. It is also of interest to determine the effect of inlet temperature with moisture content in the range representative of flue gas derived from burning lignite. Increased flue gas moisture content and decreased spray absorber inlet temperature were simultaneously accomplished by cooling the flue gas entering the spray absorber by adiabatic humidification using spray nozzles. Figure 3 summarizes the test results. Flue gas originally at 177°C (350°F) and a moisture content of 10 percent by volume will, after adiabatic humidification to 149°C (300°F), contain moisture at 11.5 percent by volume. As in the case of the heat exchanged gas, the water required in the absorber to reach the same approach to saturation will be lower when the inlet gas has already been partially prequenched. The lower water requirements and the corresponding decrease in the initial droplet surface area are reflected in the decreasing SO_2 removal efficiency with decreasing temperature (increasing extent of humidification) seen in Figure 3. When a gas is cooled at constant absolute humidity (by surface heat exchange), its adiabatic saturation temperature decreases. When it is cooled by adiabatic humidification (by prequench), its adiabatic saturation temperature remains the same, but its absolute humidity increases. As a result, for the same inlet gas temperature, more water is required by the flue gas cooled by surface heat exchange than by the prequenched (humidified) flue gas to reach the same approach to saturation. Based on this difference in water requirement, the SO_2 removal efficiency for the heat exchanged gas would be expected to be higher than that for the prequenched gas for the same inlet gas temperature. The experimental results shown in Figures 2(a) and 3(a) confirm this theoretical expectation. Moreover, the rate of decrease of the spray absorber efficiency with inlet temperature for the prequenched gas is about triple that for the precooled gas. At the same time, the SO_2 removal efficiency of the baghouse is correspondingly greater for the prequenched gas than for the precooled gas at the same inlet temperature.

STRAIGHT-THROUGH AND WET RECYCLE TESTS

Comparison with Theoretical Model

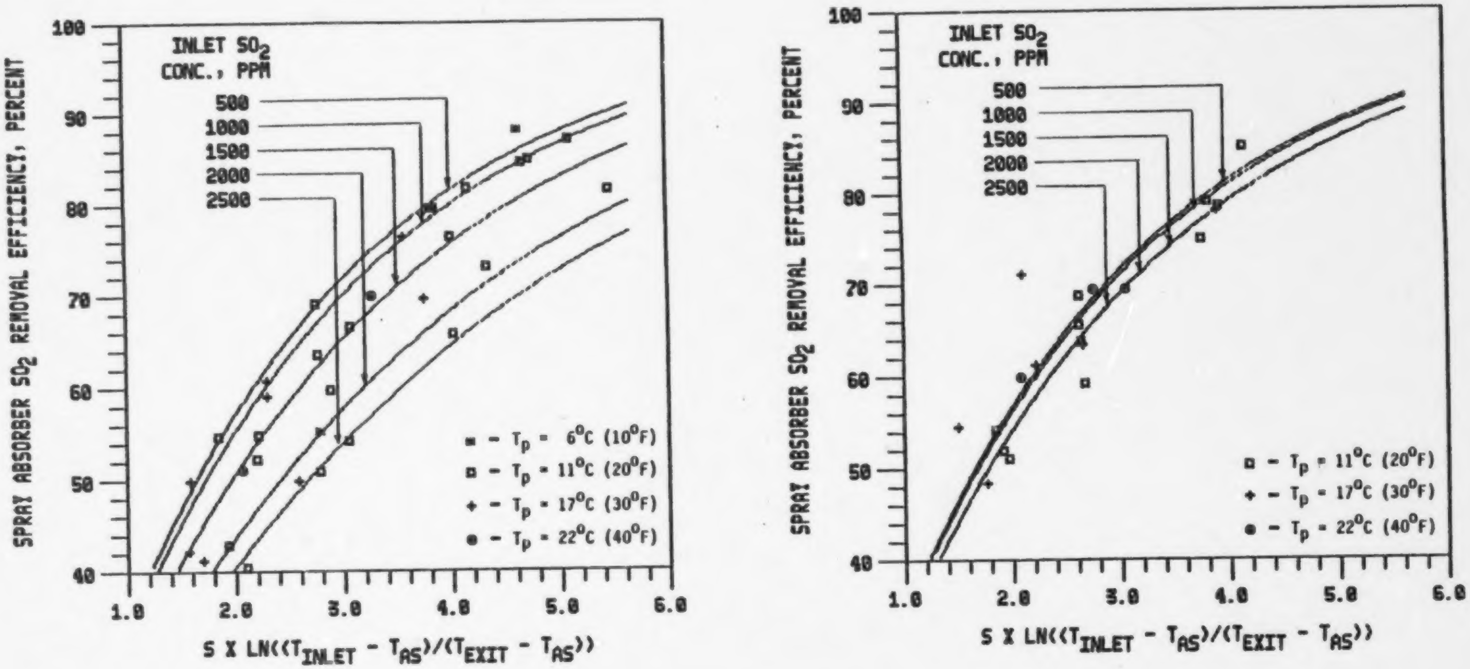
Parametric tests were performed to characterize SO_2 removal in the spray absorber/baghouse system with respect to variations in stoichiometric ratio (0.7 to 1.5), approach to saturation (7 to 22°C, 12 to 40°F), inlet SO_2 concentration (500 to 4,000 ppm), and recycle flyash (mixtures of flyash derived from eastern, western, and lignitic coals with the Martin Drake flyash). Other spray absorber operational parameters [such as atomizer angular speed (12,500 rpm), inlet gas flow rate (4.01 m^3/s, 8,500 cfm), and inlet gas temperature (177°C, 350°F)] were fixed for these parametric tests. Figure 4 presents the results of the parametric tests. When the chemical reaction between the lime and SO_2 is the rate limiting step, the theoretical model (Equation 1) predicts that all points belonging to a fixed inlet concentration should correlate on the same curve. Figure 4 shows good agreement (as supported by correlation coefficients better than 0.9) between measurements and the theoretical model predictions. The measurements and the fitted curves show a concentration-dependent performance in the spray absorber, with the lowest concentration (500 ppm) giving



3(a). Dependence of straight-through SO_2 removal efficiency on the spray absorber inlet gas temperature varied by incomplete adiabatic humidification of the inlet gas (prequench).

3(b). Dependence of the SO_2 removal efficiency under wet recycle conditions on the spray absorber inlet gas temperature varied by incomplete adiabatic humidification of the inlet gas (prequench).

Figure 3. SO_2 removal efficiency in pilot lime spray absorber/baghouse system for both straight-through and wet recycle of offproduct operation with water prequenching of the flue gas ahead of the spray absorber. [For these tests, the following parameters were fixed: inlet SO_2 concentration (1000 ppm), stoichiometric ratio (1.0), average gas residence time in spray absorber (10 seconds), atomizer angular speed (12,500 rpm), inlet gas disperser configuration (20° vane angle with vertical and small vane inserts), and approach to saturation (11°C, 20°F).]



4(a). Spray absorber SO_2 removal efficiency data for straight-through operation in the absence of external flyash injection compared with Equation 1. (regression coefficients for the correlations were approximately 0.9.)

4(b). Spray absorber SO_2 removal efficiency data for operation with wet recycle of offproduct in the absence of external flyash injection compared with Equation 1.

Figure 4. Comparison of spray absorber SO_2 removal efficiency data (points) from pilot lime spray absorber/baghouse system with predictions of Equation 1 (curves) for straight-through and wet recycle operation.

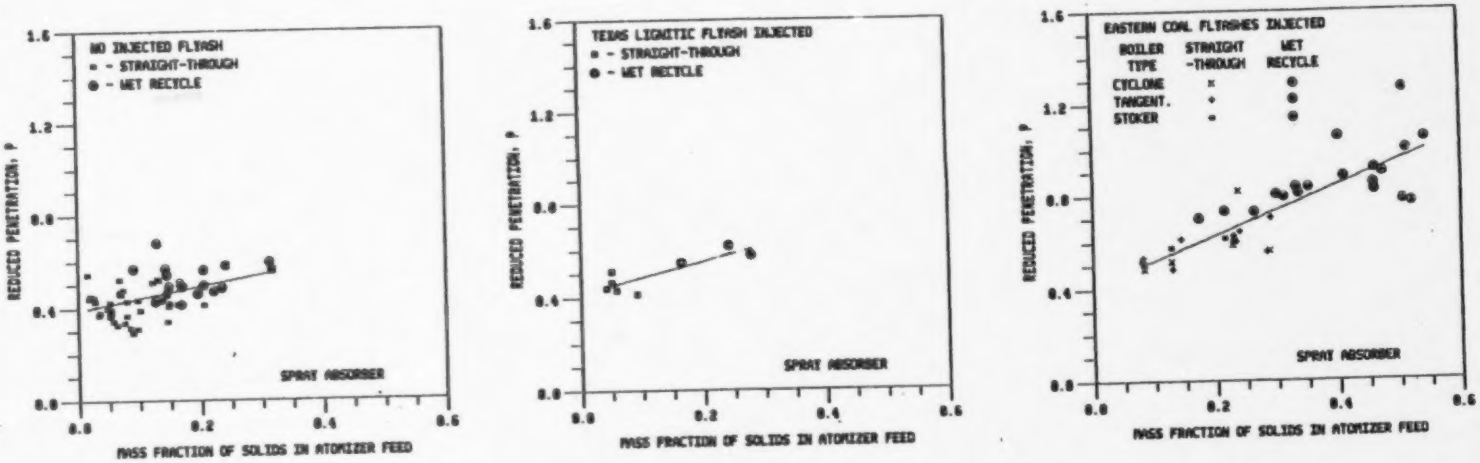
the best performance. Successive curves in Figure 4 are separated by a change in the inlet concentration of 500 ppm. For a fixed inlet gas temperature and inlet gas moisture content, the rate of water flow in the atomizer (rate of water evaporated in the spray absorber) is fixed by the approach to saturation at the spray absorber outlet. At the same time, the lime feed rate to the atomizer required to maintain a fixed stoichiometric ratio increases with increasing inlet SO_2 concentration. Consequently, the fraction of solids (lime) in the atomizer feed (or in a newly produced slurry droplet) increases with increasing inlet SO_2 concentration for a fixed stoichiometric ratio and approach to saturation. The observed deterioration of the spray absorber SO_2 removal efficiency under straight-through conditions with increasing inlet SO_2 concentration in the range of 1000 to 2500 ppm may be explained by the decreasing evaporation time of a slurry droplet with increasing SO_2 concentration. Figure 4 shows the straight-through SO_2 removal efficiency in the spray absorber to be nearly independent of inlet SO_2 concentration in the range 0-1000 ppm. This concentration-independent performance in the spray absorber for low inlet SO_2 concentrations is also predictable by the analytical model. In this case, the solids fraction in the atomizer is low enough that it essentially has no effect on the droplet evaporation time. Figure 4 also shows that the introduction of partial wet recycle of the offproduct results in the near restoration of concentration-independent spray absorber SO_2 removal. It appears that this recycle enhancement results from improved reaction sites and a greater lime availability at the same stoichiometric ratio.

Characterization of the Flyash Effect

The correlation based on Equation 2 for different types of flyash in the slurry droplets is illustrated in Figure 5. The pertinent ash characteristics are presented in Table 3. The notable feature of the correlation is the inclusion of straight-through as well as wet recycle data for all combinations of SO_2 concentration, stoichiometric ratio, and approach to saturation on a single curve. It is equally noteworthy that, despite physical and chemical differences, the data involving all three types of eastern flyash correlate on a single curve.

The recycle enhancement of SO_2 removal efficiency is the lowest for the Martin Drake flyash (no external flyash injection) which gives a nearly neutral hydrolysis (pH ≃ 7). The recycle enhancement is higher in the presence of the flyash derived from the Texas lignite which leads to an alkaline hydrolysis (pH ≃ 8). The performance enhancement from wet recycle is dramatic in the presence of flyash derived from eastern coal which yields acidic hydrolysis (pH ≃ 4). This extremely high recycle enhancement supports application of the DFGD technique to high sulfur coal.

Under straight-through conditions, the effect of an increased solids fraction was to decrease SO_2 removal efficiency in the spray absorber due to the decrease in the evaporation time of the slurry droplets. In contrast, under wet recycle conditions, the effect of an increased solids fraction appears to be an increased spray absorber SO_2 removal efficiency. This increased efficiency suggests that the beneficial effects of flyash in



5(a). Reduced penetration in the spray absorber with no injected flyash compared with Equation 2.

5(b). Reduced penetration in the spray absorber with injection of Texas lignitic flyash compared with Equation 2.

5(c). Reduced penetration in the spray absorber with injection of various eastern coal flyashes compared with Equation 2.

Figure 5. Comparison of spray absorber reduced penetration data (points) from pilot lime spray absorber/baghouse system with predictions of Equation 2 (straight lines) for straight-through and wet recycle operation with and without external flyash injection.

Table 3. CHEMICAL AND PHYSICAL CHARACTERISTICS OF FLYASH EVALUATED FOR THEIR EFFECT ON SO_2 REMOVAL IN PILOT SPRAY DRYER

	Composition, percent by mass				
Constituent or Characteristic	Stoker [a] Eastern Coal	Tangential [b] Eastern Coal	Cyclone [c] Eastern Coal	Front-fired [d] Western Coal	Lignite[e] flyash
SiO_2	34.00	45.77	27.99	62.44	60.05
Al_2O_3	14.34	19.26	13.47	23.17	21.91
CaO	2.19	2.05	1.49	3.15	8.87
Fe_2O_3	23.64	25.96	21.01	3.70	3.09
Na_2O	0.49	0.41	0.31	0.21	0.46
K_2O	1.90	1.98	1.75	1.20	0.89
Li_2O	0.02	0.02	0.02	0.02	0.02
MgO	1.17	0.90	0.69	1.27	2.04
P_2O_5	0.38	0.16	0.24	1.22	0.10
SO_3	1.11	0.45	0.53	0.45	0.31
LOI[f]	19.40	1.42	31.34	1.85	0.10
Initial pH[g]	4.7	4.4	3.2	>7.0	>7.0
Alkalinity, meq/g[h]	-0.062	-0.038	-0.307	0.180	5.12
Mass median diameter, μm	67	28.5	14.3	17.3	—
Log-normal standard deviation[i]	4.1	4.3	2.6	2.8 2.8	—
Volume-to-surface diameter, μm[j]	24.8	9.8	9.1	10.2	—
Specific surface area, m^2/kg	106	251	253	280	—
Typical surface area at spray dryer inlet m^2/m^3 (ft^2/ft^3) gas	0.62(0.19)	1.15(0.35)	0.59(0.18)	1.15(0.35)	—
Typical concentration at spray dryer inlet: 10^{-3} kg/m^3 (gr/ft^3)	5.7(2.5)	4.6(2.0)	2.3(1.0)	4.1(1.8)	8.0(3.5)

[a] Stoker-fired boiler at the University of Iowa

[b] Tangentially fired boiler in Units 5 and 6 of the Conesville Station operated by Columbus and Southern Ohio Electric

[c] Cyclone-fired boilers in Unit 2 of the Conesville Station operated by Columbus and Southern Ohio Electric (inlet hopper sample)

[d] Front-fired boiler in Unit 6 of the Martin Drake Station operated by the City of Colorado Springs

[e] Flyash derived from burning Texas lignite at the Monticello Station operated by Texas Utilities Services

[f] LOI stands for loss on ignition at 800°C (1472°F).

[g] Initial pH of 0.01 kg fly ash in 0.1 ℓ water at 52°C (126°F)

[h] Negative alkalinity indicates acidity being titrated against base to a final pH of 7.5.

[i] Log-normal standard deviation, σ, is obtained from $\sigma = 0.5 (z_1/\bar{z} + \bar{z}/z_2)$ where 15.9, 50.0, and 84.1 percent of the mass of the particle distribution originate from particles whose diameters are greater than z_1, $\bar{z}$, and z_2.

[j] Equal to $d_m \exp[-\frac{1}{2}(\ln \sigma)^2]$ where d_m is the mass median diameter and σ is the log-normal standard deviation

enhancing the rate of chemical reaction between the SO_2 and lime override the limitation in efficiency due to reduced evaporation time.

CONCLUSION

The data obtained from the Colorado Springs DFGD test program strongly support the models given by Equations 1 and 2. As a result, SO_2 removal performance can now be accurately predicted for a conservatively designed system when the stoichiometric ratio, approach to saturation temperature, and slurry solids content are defined. The predictions can be extended to systems using ash recycle when an additional empirical constant derived from the flyash type is available.

An important DFGD feature which was identified is the manner in which the spray absorber and baghouse work in unison toward maximizing the system SO_2 removal efficiency. This trend is evident in the variation of the SO_2 removal efficiencies with inlet temperature. The decrease in spray absorber performance with decreasing inlet temperature is seen to be compensated by an increase in the baghouse performance to result in a nearly constant system removal. Conditions of high humidity and low inlet temperature are favorable to good SO_2 removal in the baghouse. Conversely, conditions of low humidity and high inlet temperature appear to be favorable to good SO_2 removal by the spray absorber. Significantly, however, the system performance appears to be only weakly dependent on either the inlet temperature or the humidity which may occur either as a result of variations in coal moisture content or variations in the ambient (combustion) air. The spray absorber/baghouse system appears capable of effecting good performance even in the presence of these variations when no baghouse reheat is provided. Baghouse reheat removes the compensating character of baghouse operation and allows the spray absorber sensitivity to inlet humidity and inlet temperature to predominate.

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DRY SCRUBBER, FLUE GAS DESULFURIZATION ON HIGH-SULFUR, COAL-FIRED STEAM GENERATORS: PILOT-SCALE EVALUATION

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ABSTRACT

This paper describes the pilot-scale investigation of methods for adapting dry flue gas desulfurization (FGD) to utility steam generators burning high-sulfur coal. Development of the dry scrubber for FGD was initially directed toward reducing SO_2 in steam generators burning low-sulfur, western coals. The reason for limiting dry scrubbing to low-sulfur coals was two-fold: First, federal New Source Performance Standards (70% reduction) were less stringent. Second, western coals generally contain less sulfur and large amounts of ash alkali. This significantly contributes to the effectiveness of the dry scrubbing process. Several potential drawbacks (both technical and economic) limiting dry scrubber technology to western coals are discussed.

The Department of Energy (DOE) conducted dry scrubber FGD tests in 1981. Preliminary results indicated dry scrubbing could remove more than 90% of the sulfur released from the combustion of eastern coals at less-than-anticipated levels of lime consumption. The Babcock & Wilcox Company (B&W), under DOE contract, has tested the dry scrubbing process to evaluate the variables that have a major effect on SO_2 capture.

There are potential ways of increasing SO_2 capture via dry scrubbing. One method requires limestone injection into the boiler furnace, reducing SO_2 levels to the scrubber, calcines limestone to more reactive lime, and simulates a high-alkali-ash, high-sulfur coal. Other methods covered include recycle and dry scrubber approach to saturation at temperatures less than 20°F to obtain high SO_2 capture.

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INTRODUCTION

The use of dry scrubbers for controlling SO_2 emissions from power plants burning low-sulfur coal has gained wide industrial acceptance. A natural outgrowth of the success of this technology is to establish its limits of application in regard to higher-sulfur coals. The interest in developing dry scrubber technology has been accelerated due to growing concerns about controlling acid precipitation.

BACKGROUND

Using dry scrubbers for flue gas desulfurization began in the late 1970s. The first commercial utility units are operating on low-sulfur coal. B&W's development of dry scrubber technology has been a four step process:

- A small, field, pilot plant (8000 ACFM) was constructed and put into operation in 1978 at the Basin Electric Neal Station, Velva, N. D. This pilot provided basic process information but limited information for scale-up and system sensitivity to fuel changes.
- To satisfy the need for additional design data, B&W proceeded with construction and operation of a 120,000-ACFM field demonstration unit to demonstrate mechanical scale-up of the system.
- B&W then constructed a smaller 1500-ACFM pilot plant at the its Alliance (Ohio) Research Center (ARC). This unit was designed to develop basic process data.
- The net result of B&W's development efforts has been construction and start-up a dry scrubbing system consisting of four units of 800,000 ACFM each (600 MW total) at Basin Electric's Laramie River Station, Wheatland, Wyoming. (Another 450-MW unit at Colorado Ute's Craig Station is scheduled for start-up in 1984.)

The data obtained from the B&W pilot plants and demonstration unit has been used to establish a basis for predicting performance for dry scrubber systems.

PROGRAM OBJECTIVES

The objective of the B&W-DOE program was to evaluate dry scrubber technology when applied to high-sulfur coals. The first step was to develop parametric baseline data using standard dry scrubber practices. This would be followed by using innovated methods for improving performance. This program was funded within the DOE Advanced Environmental Control Technology (AECT) plan initiated in 1979. The AECT program will develop the technology base for controlling contaminants produced during coal conversion [1].

To accomplish the objectives, a wide range of process data were obtained. We decided that the B&W pilot plant at (ARC) would be most suitable to meet these objectives. This unit was selected because it was small enough (1500 ACFM) to keep operating costs reasonable. Also, the unit was characterized against a larger (120,000-ACFM) field demonstration unit that provided a sound basis for scaling up the data.

DESCRIPTION OF TEST FACILITIES

B&W's 1500-ACFM, pilot dry scrubber (Figure 1) is highly flexible and designed so a wide range of variables could be investigated. The system starts with a combustion chamber that has a rating of 5×10^6 Btu/Hr heat input. The combustor is also capable of firing oil or gas. The exhaust gas from the combustor can be directly vented to the atmosphere or diverted to the dry scrubber pilot facility. The gas stream going to the scrubber can be conditioned with fly ash and/or SO_2. This capability provides a way of simulating flue gas pollutants without burning actual fuel. The flue gas can travel directly to the dry scrubber or pass through a heat exchanger that controls the flue gas temperature at the scrubber inlet.

The first compartment of the dry scrubber is shown in Figure 2. A perforated plate is located at the inlet for distribution purposes. The gas then passes through a TurboDiffuser™, where swirl energy is introduced to the gas stream. The gas is then intimately mixed with a finely atomized reagent slurry as it leaves the TurboDiffuser™. Here, the SO_2 reacts with the spray's alkali as moisture in the spray is evaporated into the flue gas. A portion of dry, spent product consisting of fly ash, calcium sulfates, calcium sulfites, and unreacted lime deposits in the hoppers at the bottom of the scrubber (Figure 3). The remainder is carried to a cyclone collector and/or baghouse, where the majority of the particulate by-product is removed. The gas stream is then exhausted through an induced-draft fan to the atmosphere.

The lime slurry reagent preparation system shown in Figure 1 is composed of a reagent hopper feeding dry bulk reagent through a rotary valve to a commercial paste slaker equipped with an automated grit separator. The lime putty then discharges by gravity through a screen to a heated slurry tank. The tank is continuously agitated and the slurry is continuously pumped around a recycle loop to avoid potential settling out of the reagent and associated plugging problems. The slurry feed line to the dry scrubber is equipped with a strainer to protect the spray nozzle from plugging. Feed slurry is introduced to the scrubber through a pneumatic spray nozzle (Figure 4).

TEST PLAN

A test plan was developed based on dry scrubber data obtained during earlier development by B&W on low-sulfur coals. The starting point was to establish a list of process and operating variables that were known to influence process performance. (These variables are listed in order of significance on Table 1; the test plan is shown on Table 2.) Included in the test plan were methods for improving the overall process performance for dry scrubbers when used in high-sulfur coal applications; these methods are:

- Use of recycle.
- Use of low temperature at the scrubber outlet.
- Use of furnace limestone injection in combination with the dry scrubber.

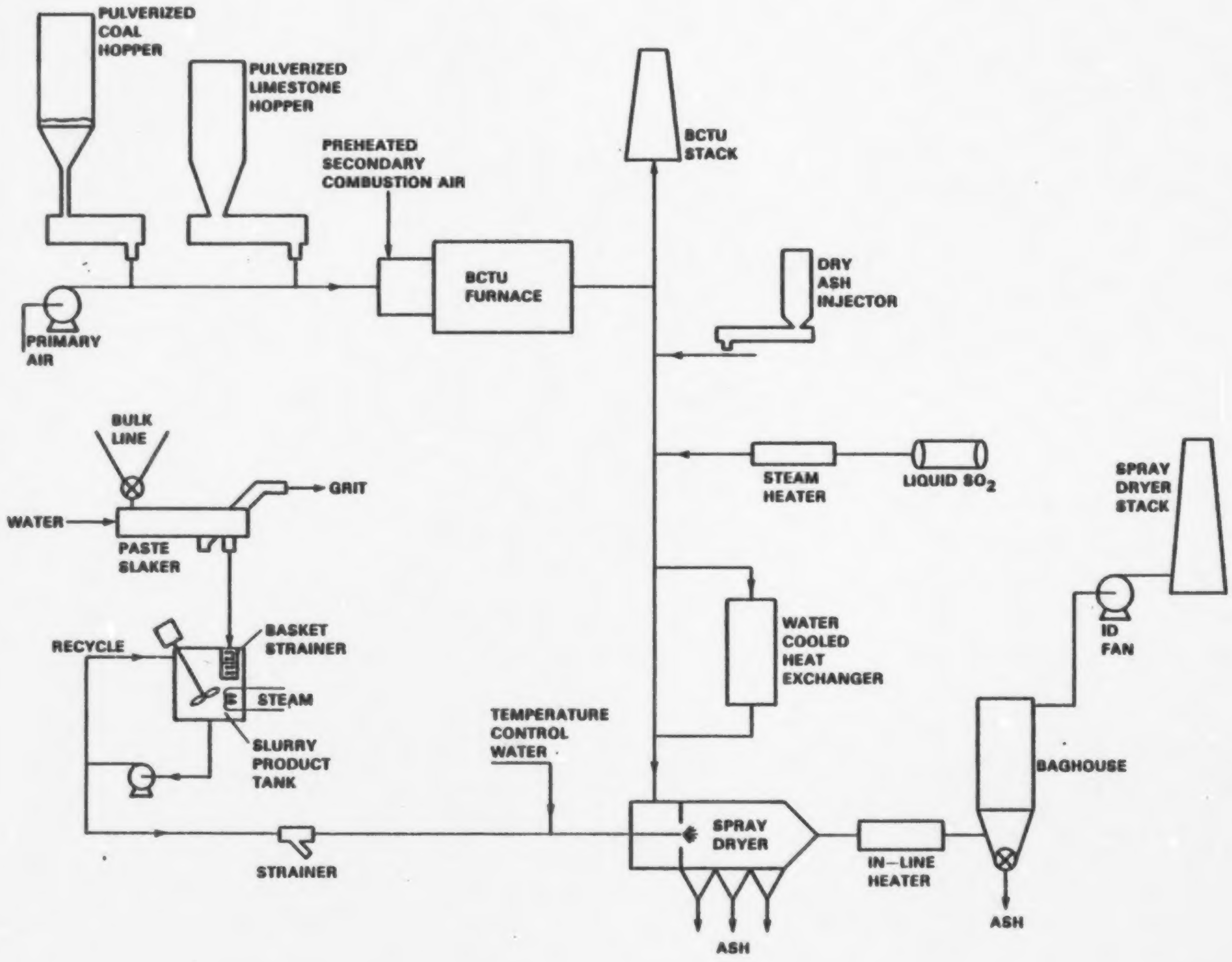
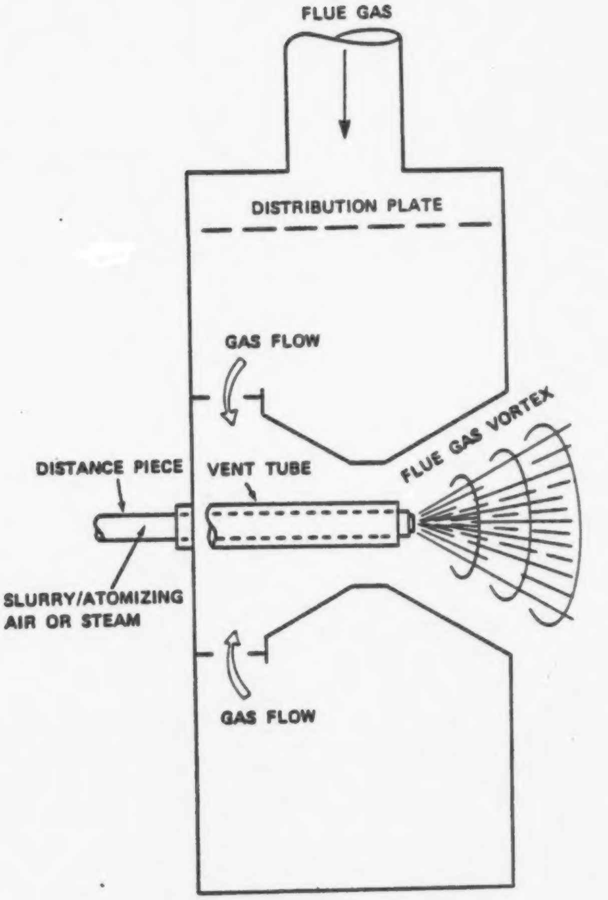


Figure 1. Babcock & Wilcox 1500-ACFM dry scrubber pilot.



*VENT TUBE CLOSED DURING ARC TESTING

Figure 2. Dry scrubber plenum with distributor plate and throat configuration.

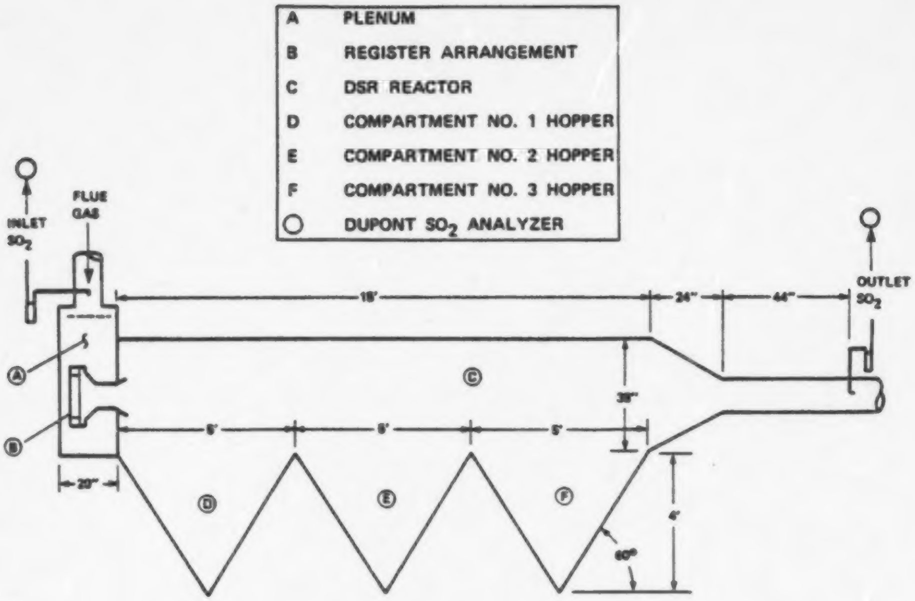


Figure 3. Dry scrubber dimensions.

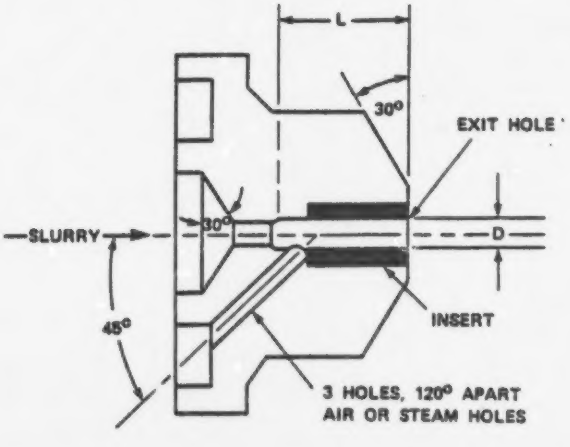


Figure 4. Advanced sprayer plate for pneumatic atomization of slurries.

Table 1

PROCESS AND OPERATING VARIABLES INFLUENCING DRY SCRUBBER SYSTEM PERFORMANCE

Process	Operating
• Reagent type	• Stoichiometry
• Dryer inlet temperature	• Dryer outlet temperature
• Dryer inlet SO_2 concentration	• Solid recycle rate
• Fly ash composition	• Dryer residence time

Table 2

DESCRIPTION OF DRY SCRUBBER TEST VARIABLES

Baseline tests:

  - Stoichiometric ratio
  - Dryer inlet temperature
  - Dryer outlet temperature
  - Baghouse outlet temperature
  - Baghouse air-to-cloth
  - Baghouse pressure drop
  - Dryer residence time
  - Reagent variation
  - Coal variation

Recycle tests

Low-temperature outlet tests

Limestone injection

PILOT PLANT TEST RESULTS

The major emphasis of dry scrubber testing was to evaluate utilizing lime slurry reagent at high concentrations of SO_2 at the inlet under a variety of process and operating conditions. Holding other variables constant, testing began by generating a nonrecycle data base that included a variation of both reagent and coal. The first series of tests were designed to show the effect of stoichiometry (over a wide range) on SO_2 capture. Subsequent test series looked at recycle, low temperature at the scrubber outlet, and limestone injection as an alternative means of improving lime utilization. Baghouse, dryer residence time, and coal variation test results are not yet available.

Lime slurry reagent was prepared on site with a paste slaker. The Mississippi pebble lime specifications are listed in Table 3. Typical chemical analysis of pebble lime samples taken during slaking operation are shown in Table 4. The lime slurry has approximately 50% wt/wt below 7-micron-diameter particles.

The test furnace was fired with a Western Kentucky bituminous coal; chemical analyses of the coal and ash are shown in Tables 5 and 6. The sulfur content of this coal ranged from 2.6% to 3.0%; furnace excess air ranged from 15% to 20%.

STOICHIOMETRIC RATIO AND OUTLET TEMPERATURE VARIATION

Once-through dry scrubber plus baghouse SO_2 absorption values are shown in Figure 5; typical inlet conditions were:

Inlet SO_2 level	—	1900 - 2300 ppm
Inlet Gas Temperature	—	290° - 311°F
Dryer Gas Residence Time	—	9.6 to 10.8 sec
Atomizing Air Consumption	—	0.17 to 0.21 lbs air/lbs slurry

Baghouse inlet temperature was held at least 20°F above the flue gas adiabatic saturation temperature to prevent bag blinding. The approach to saturation temperature (AST) is calculated as the difference between the flue gas bulk temperature and its adiabatic saturation temperature.

The significance the dry scrubber's outlet temperature has on SO_2 capture is evident in Figure 5. At a stoichiometry of 1.15, approximately 90% SO_2 capture was measured at a 10°F AST.

Dry scrubber absorbers have been recently tested on high-sulfur coals by several researchers. At a lime feed rate corresponding to a stoichiometric ratio of 1.2, and inlet SO_2 concentration of 2000 PPM, Samuel, et al. [2], obtained 84% SO_2 capture with an AST of 20°F. Yeh, et al. [3], using commercial hydrated lime, obtained approximately 65% SO_2 capture at a 1.0 stoichiometry and inlet SO_2 concentration of 2100 ppm and a 30°F AST. These results are consistent with the baseline data in Figure 5.

Table 3

TYPICAL MISSISSIPPI PEBBLE LIME SPECIFICATIONS

Mississippi or Peerless Rotary Plant

Pebble Quickline

Size Designation 1/2 - 1 - 2 inches

Chemical Analysis

CaO — available	93.0%	to	97.0%
CaO — Total	96.0	to	98.0
$CaCO_3$	0.3	to	4.0
$CaSO_4$	0.04	to	0.21
S — equivalent	0.01	to	0.05
SiO_2	0.75	to	1.10
Al_2O_3	0.15	to	0.35
Fe_2O_3	0.08	to	0.12
MgO	0.53	to	0.75
LOI	0.30	to	2.00

Table 4

CHEMICAL ANALYSIS OF MISSISSIPPI PEBBLE LIME

Date	022682	042382
Moisture (%)	Negligible	Negligible
S as SO_3 (%)	0.1	0.06
CO_3 as CO_2 (%)	1.0	0.17
Ca as CaO (%)	93.3	96.5
Mg as Mg (%)	0.6	not available

Table 5

TYPICAL CHEMICAL ANALYSIS: WESTERN KENTUCKY COAL

Dry Scrubber
ACG-83-4030-31
April 5, 1982

Sample No.	C-16459		C-16460	
Description	Pulverized Coal Run 0011, 031082 0700		Pulverized Coal 031282, 0830	
Basis	As Received	Dry	As Received	Dry
Total Moisture, %	3.8	—	3.8	—
Proximate Analysis, %				
Moisture	3.8	—	3.8	—
Volatile Matter	40.9	42.5	41.6	43.3
Fixed Carbon	45.1	46.9	44.4	46.1
Ash	10.2	10.6	10.2	10.6
Gross Heating Value, Btu per lb.	12450	12940	12420	12910
Btu per lb. (M&A-Free)	—	14470	—	14450
Ultimate Analysis, %				
Moisture	3.8	—	3.8	—
Carbon	69.1	71.9	68.8	71.5
Hydrogen	4.9	5.1	4.8	5.0
Nitrogen	1.33	1.38	1.38	1.43
Sulfur	2.92	3.04	3.05	3.17
Ash	10.2	10.6	10.2	10.6
Oxygen (Difference)	7.75	7.98	7.97	8.30
Total	100.00	100.00	100.00	100.00

Table 6

TYPICAL CHEMICAL ANALYSIS: WESTERN KENTUCKY COAL ASH

Dry Scrubber
ACG-83-4030-31
April 5, 1982

Sample No.	C-16459	C-16460
Description	Pulverized Coal Run 0011, 031082 0700	Pulverized Coal 031282, 0830
Ash Analysis (Spectrographic), %		
Silicon as SiO_2	48.61	48.86
Aluminum as Al_2O_3	18.75	19.26
Iron as Fe_2O_3	19.26	19.92
Titanium as TiO_2	0.99	0.99
Calcium as CaO	3.04	3.12
Magnesium as MgO	0.92	0.90
Sodium as Na_2O*	0.51	0.53
Potassium as K_2O*	2.83	2.71
Sulfur as SO_3	2.72	3.05
Phosphorus as P_2O_5	0.42	0.44

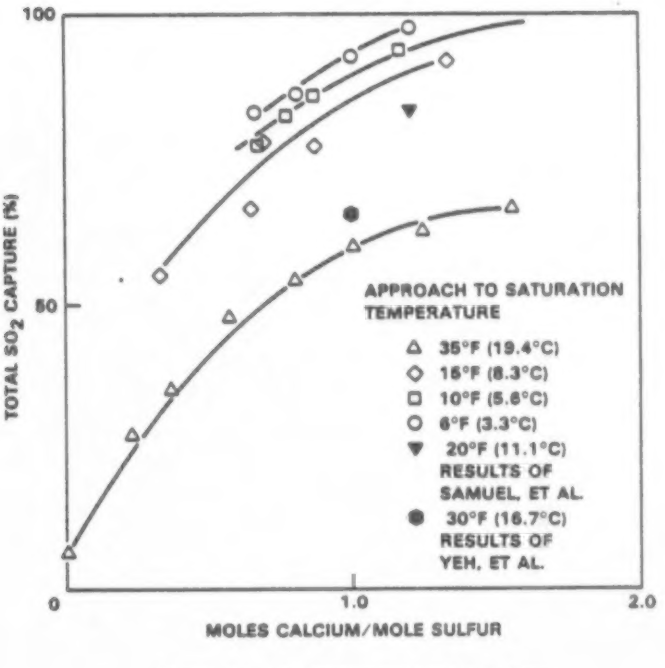


Figure 5. Stoichiometric ratio testing.

In general, eastern coals contain relatively small amounts of low-alkaline ash when compared to Western coals. Previous B&W work on several western coals indicated dry scrubber SO_2 capture of 25% - 35% at 0.0 stoichiometry. For the 2000 ppm SO_2 flue gas treated at a 35°F AST, total SO_2 capture was less than 10%, as expected, at a 0.0 stoichiometry.

TEMPERATURE VARIATION AT SCRUBBER INLET

Evaporation time of slurry droplets containing relatively insoluble solids is characterized by the slurries' physical properties (solids content, density, initial particle diameter, and temperature) and the dryer inlet and outlet temperature. Typical dry scrubbers in FGD applications reduce inlet temperature by 100° - 150°F (38° - 66°C). The effect of this temperature difference on SO_2 capture for western coals was shown to be very small by Parsons, et al. [4]. Figure 6 shows the effect of increasing dry scrubber inlet temperature for a 35°F AST and a 1.0 stoichiometry. This indirectly shows how scrubber temperature drop affects SO_2 capture and clearly indicates a significant influence of inlet temperature on SO_2 capture. Previous B&W tests with high-sulfur coal also implied inlet temperature can play a more dominant role on SO_2 removal for high-sulfur coals in dry scrubber applications [5].

REAGENT VARIATION

Commercial and industrial dry scrubber FGD systems have relied on lime slurries as the principal reagent. Limestone slurries have shown acceptable SO_2 absorption properties in low-sulfur, wet scrubber FGD. But in dry scrubbing FGD, these slurries have unacceptable reactivity. Research in various lime and, most recently, sodium-based reagents (such as Trona [4] and soda ash [3]) has characterized relative reactivities.

Johnson [6] performed tests with fly ash alkali and concluded certain high-calcium ash, western coals contained enough inherent alkali to obtain the required SO_2 capture. The majority of commercial dry scrubber systems today use lime reagents; one exception is Rockwell's Coyote Station dry scrubber, which uses soda ash.

Table 7 shows chemical analysis of two reagents tested for SO_2 capture. The soda ash was a granular, high-quality (99+%), commercial-grade sodium carbonate (Na_2CO_3). This material is a much stronger base than lime and forms hygroscopic solids, both properties enhancing SO_2 capture in the aqueous- and dry-phase chemistry. A thiosorbic lime containing 4.8% Mg as MgO was also tested. Figure 7 indicates total SO_2 capture for soda ash, thiosorbic lime, and the base Mississippi lime. As expected, the soda ash outperformed both lime reagents. The thiosorbic lime tests indicate poorer SO_2 capture relative to the base lime. As noted, the thiosorbic lime supplier reported excess carbonate core. This was due to low kiln temperatures. While the reported slaking time is well within the 3-minute-limit slaking time for high-reactivity lime, these tests should be repeated.

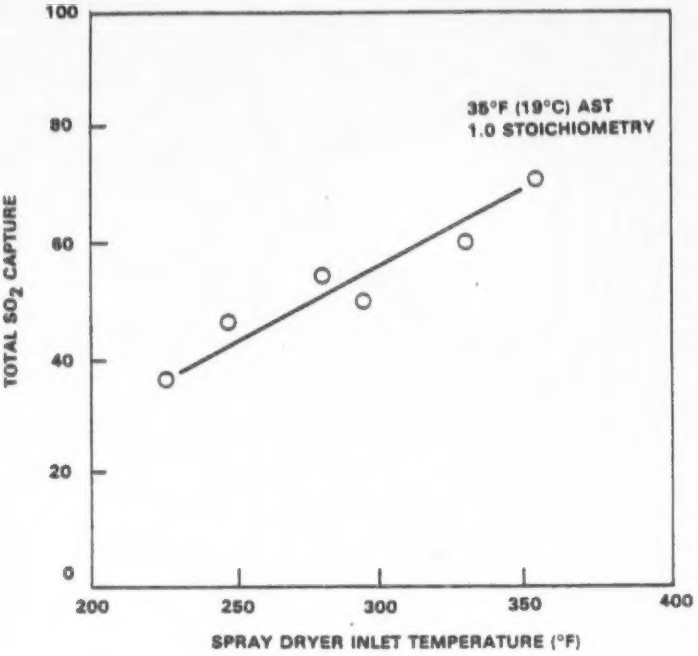


Figure 6. Effect of gas inlet temperature on SO_2 capture.

Table 7

REAGENT VARIATION CHEMICAL ANALYSIS

Sample No.	M-38236	M-38237
Description	Soda Ash 7-6-83 1500	Thiosorbic Lime 7-5-83
Basic	As-Received	As-Received
Calcium as CaO, %	—	88.35
Magnesium as MgO, %	—	4.85
Carbonate as CO_2, %	41.46	0.39
Total Sulfur as SO_3, %	—	0.34
Total Insolubles, % (including SiO_2)	—	3.00
Carbonate as Na_2CO_3, % (as-recvd., calculated)	99.84	—

Note: Vendor reported thiosorbic shipment contained 3.1% excessive core, 91.9% available lime index, and 1-minute slaking time.

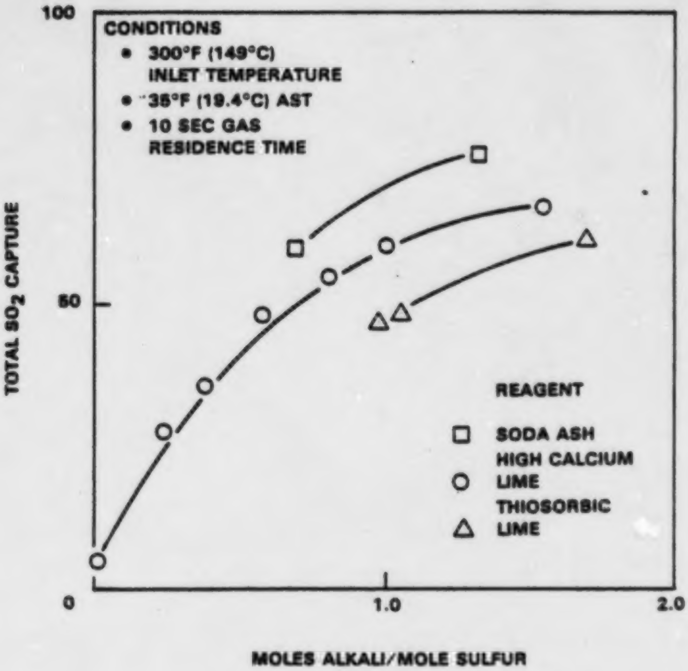
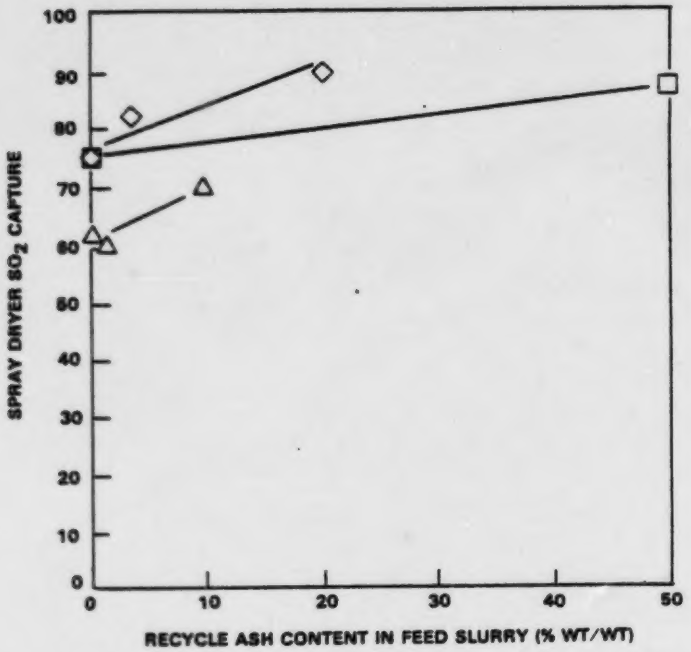


Figure 7. Reagent variation testing.

RECYCLE TESTS

The degree of reagent utilization in a dry scrubber system has a significant effect on system economics. One significant difference between wet and dry scrubbing systems is the lower reagent utilization in dry scrubbers. This is mainly due to type and rate of gas-liquid contact. Wet scrubbers use relatively large liquid-to-gas ratios, on the order of 40 - 90 gpm/1000 ft^3 gas, treated (53 - 119 L/M^3) to achieve intimate gas-liquid contact with counter-current absorption trays. Dry scrubbers in FGD have average gas-liquid ratios of 0.2 - 0.3 gpm/1000 ACFM (0.267 - 0.40 L/M^3) and generally use co-current gas-liquid contact via a gas diffuser and atomizer.

Recycle of solid material collected in the dry scrubber and particulate collector can improve reagent utilization by reusing unreacted calcium hydroxide and other fly ash alkali. The second source of alkaline material — fly ash — originates from calcined carbonates of calcium, sodium, and potassium formed during coal combustion. Figure 8 shows the increase in dry scrubber SO_2 capture when the feed slurry ash content increases for a 33°F (18°C) and 19°F (10.6°C) AST. Approximately 7% more sulfur was captured at a 10% wt/wt wet recycle rate. Samuel, et al. [2] conducted similar tests at substantially higher recycle rates. Note that his work was at a 1.2 stoichiometry and a utilization at no recycle of 63%.



	TEST RUNS	Ca/S	AST (F/C)	INLET SO_2 (PPM)	INLET TEMP. (F/C)	SOURCE
△	BASE, 175, 176	1.6	33/18	2200	303/151	B&W
◇	BASE, 183, 184	1.6	19/10.6	2150	305/152	B&W
□	—	1.2	20/11.1	2000	—	REF 4

Figure 8. Wet recycle spray of dry scrubber system solids.

Table 8

CHEMICAL ANALYSIS OF RECYCLE TEST SLURRYS

Description	Solids, %	Calcium as CaO, % Dry	Carbonate as CO_2, % Dry	Sulfur as SO_3, % Dry
Base Lime Tests 175 & 176	23.9	71.1	0.29	0.04
Test 175	25.9	67.2	3.42	4.05
Test 176	24.2	69.0	2.07	2.36
Base Lime Tests 183 & 184	19.6	72.4	0.46	0.05
Test 183	23.0	62.6	6.52	9.96
Test 184	20.2	65.6	4.90	4.46

Test Runs 183 and 184 were at the same AST but, due to the higher stoichiometry, attained only 47% utilization. The difference in marginal increase of SO_2 capture based on increases in wet recycle between these two test series appears to be a function of reagent utilization with no recycle. Also, at the same SO_2 capture (no recycle), as stoichiometry increases from 1.2 to 1.6, so does the marginal effect of wet recycle on SO_2 capture. For dry scrubber FGD on high-sulfur coals, wet recycle will be essential to maintain acceptable reagent and fly ash utilization.

Table 8 lists chemical analyses of the base and recycle tests shown in Figure 8 from this program. Addition of ash to the base slurry is evident by the increase of total sulfur in forms of calcium sulfates and sulfites and carbonates, principally calcium carbonate.

LIMESTONE INJECTION

Limestone injection in a multistage burner (LIMB) uses limestone ($CaCO_3$) as reagent introduced into high-thermal-combustion zones, where calcination to lime (CaO) occurs. The solid lime then adsorbs gaseous SO_2, forming sulfites and sulfates of calcium ($CaSO_3$ and $CaSO_4$). The $CaCO_3$ dissociation reaction is principally a function of temperature and the flue gas CO_2 partial pressure. At atmospheric pressure, limestone will begin dissociation under normal coal combustion conditions at approximately 1450°F (788°C) [7]. Once the CO_2 gas has evolved, sulfation may proceed.

Obtaining a high-quality calcined limestone depends primarily on the temperature profile existing in the combustor and time profiles from the point of reagent injection. Murray [8] studied calcination efficiency on 43 commercial limestones and concluded each particular limestone has a unique time-temperature profile for optimum calcination. Dogu [9] investigated initial limestone pore structure effect on SO_2 adsorption. The average lime particle pore radius and overall porosity were measured at various calcining temperatures, with both having an optimum value at 1740°F (950°C). Dogu's studies also found diffusion resistance from pore closure by $CaSO_4$ the controlling mechanism.

Recent work to develop low-NO_x coal burners has helped limestone injection studies, since their characteristics of low flame temperature and long burn time are quite suitable for limestone injection. Dremmel, et al. [10] studied three types of coal-fired furnaces as part of EPA's simultaneous SO_x, NO_x study. Wall-, tangential-, and stoker-fired combustors were used to evaluate primarily coal firing rate, cooling rate, and sorbent injection velocity. A wall-fired furnace simulating the normal time-temperature profile seen by coal and sorbent particles was fired in a low NO_x mode. At sorbent-sulfur ratios of 2.0, 40% to 60% SO_2 capture was measured. Quantitative effects of time, temperature, and calcination rates were still undetermined.

Figure 9 lists furnace SO_2 capture versus stoichiometry for primary and secondary limestone injection (also see Table 9). At the 2000°F injection temperature, primary injection (limestone interspersed with coal) captured slightly less SO_2 than secondary injection (13.6% to 13.3%). Within the accuracy of measuring SO_2, however, these removals are essentially identical.

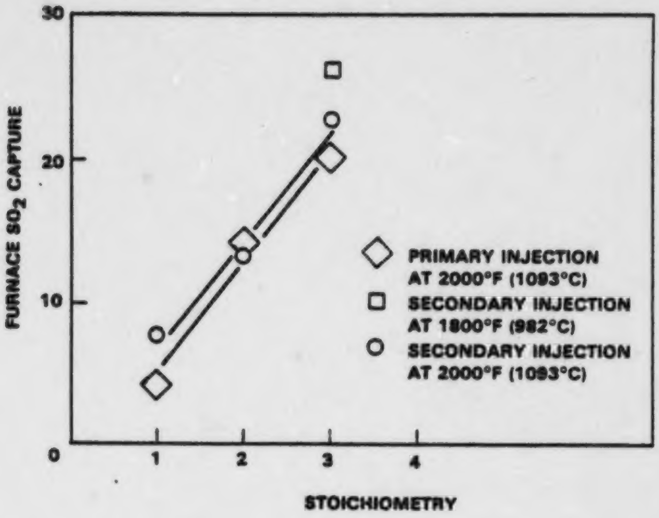


Figure 9. Limestone injection results.

Table 9

TEST RESULTS: LIMESTONE INJECTION

Injection Method	DOE Test Number	Load MBtu/hr	% O_2*	SR	Injection Temp.	% Removal in Furnace
Primary	D221	4.0	2.9	3.0	2000°F**	20.0
	D222	4.0	2.75	2.0	2000°F	13.6
	D223	4.0	2.75	1.0	2000°F	6.7
	D223	4.0	3.0	1.0	2000°F	4.1
	D223	4.0	3.13	1.0	2000°F	6.3
Secondary	D212	3.1	3.75	3.0	1800°F	26.0
	D213	2.0	3.73	2.0	1800°F	13.6
	D214***			NO TEST RESULTS		
	D215	4.0	3.17	3.0	2000°F	22.4
	D216	4.0	3.78	2.0	2000°F	13.3
	D217	4.0	3.79	1.0	2000°F	7.3

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*As measured at reactor inlet

**Furnace exit temperature

***Insufficient turndown on screw feeder

Primary reagent injection exposes the limestone particles to the highest flame temperatures and longest residence time possible. Sintering of the newly formed CaO particle pores is a major concern with this type of injection mode. For secondary injection at varying furnace outlet temperatures, a small increase was also measured in SO_2 capture as the furnace temperature was lowered. This assumes that an equilibrium between decarbonation and recarbonation had been established. Some partial pore closure by sintering is also suspected. In general, with a low-NO_x burner producing 250 - 400 ppm NO_x, approximately 25% SO_2 was captured at a 3.0 stoichiometry. The inserted limestone was pulverized to approximately 70% through 200 mesh — equivalent in size to the pulverized coal. A finer limestone grind would have increased furnace SO_2 capture by producing more solid surface area and minimizing the pore blinding effects of $CaSO_4$ formation.

DRY SCRUBBING TECHNOLOGY FOR FGD OF EASTERN COAL

Successful adaption of dry scrubber FGD systems to high-sulfur-coal applications requires first an understanding of SO_2 capture at various operating conditions meeting federal New Source Performance Standards (NSPS). Based on nonrecycle operation, these tests indicated 90% SO_2 capture can be obtained at a dry scrubber stoichiometry of 1.20 (75% utilization) and an approximate 15°F AST. At a 1.6 stoichiometry and 20°F AST, 20% solids recycle increased SO_2 capture from approximately 75% to 90%. Maintaining the higher 20°F AST required recycle and increased lime consumption to meet NSPS. However, the higher AST will reduce the possibility of a system upset. A dry scrubber treating 3500 ppm SO_2 flue gas at a 20°F AST and a 1.6 stoichiometry requires approximately 25% lime slurry. Recycle will increase slurry solids and is primarily limited by the maximum atomizer solids for acceptable atomization. For an industrial dry scrubber system, adjustments between AST, recycle rate, and stoichiometry will define actual process conditions required to meet NSPS and high availability.

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EPRI SPRAY DRYER/BAGHOUSE PILOT PLANT STATUS AND RESULTS

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ABSTRACT

In February 1982, the Electric Power Research Institute (EPRI) initiated a 2-1/2 MW spray dryer/baghouse FGD pilot plant program at their Arapahoe test facility. The objective of the pilot plant program is to confirm the capabilities of the FGD process and to provide the electric utility industry with reliable design and operating information for spray dryer/baghouse FGD systems. The pilot unit was described and initial results for sodium carbonate and once-through lime operation were presented at the May 1982 FGD symposium in Hollywood, Florida. This paper presents the results of test work conducted from May 1982 through August 1983.

The majority of the test work has been conducted with lime reagent in the recycle, rather than once-through mode. Effects of a number of variables have been studied. Spray dryer inlet SO_2 concentrations have been varied from a nominal 350 ppm up to 2000 ppm. Other variables examined have included reagent ratio, recycle rate, system flue gas flow rate, atomizer feed slurry preparation and feeding configurations, and approach to adiabatic saturation at the dryer outlet. A significant result has been the observation that recycle operation greatly improves spray dryer operation in addition to improving SO_2 removal performance.

The fabric filter has been shown to contribute significantly to overall system SO_2 removal, particularly at higher system removal levels (80 percent and greater). No bag/fabric-related problems have been observed. However, corrosion of mild steel baghouse walls and mild steel caps on bags near the walls has in some instances been severe. The corrosion has been largely attributed to insufficient insulation of baghouse surfaces, and to the fact that the pilot-scale compartment shares no common walls with other compartments. As a result of several bag cap failures, the fabric filter compartment was re-bagged in April 1983. The new bags were brought on-line with no conditioning by fly ash-only operation, and after 4 months continue to operate at a very low pressure drop.

INTRODUCTION

Currently, spray dryer based flue gas desulfurization (FGD) systems for 16 utility boilers with a total capacity of over 6000 MW have been installed or are under contract. Six of the systems are in start-up or commercial operation. However, of the two systems that have been in operation longest, one is unique in using soda ash reagent, while the other has been operated by the vendor until only recently. Thus, in spite of the large number of units sold, very little information on the design and operation of these systems is available to the utility industry.

Because of this lack of information, the Electric Power Research Institute has chosen to pilot a spray dryer/fabric filter system at their test facility adjacent to the Public Service of Colorado Arapahoe station. Numerous pilot-scale programs have been conducted by process vendors, but detailed results from these programs have generally not been published. The goals of the EPRI program are to independently assess the operability of a spray dryer and fabric filter system in FGD service, and to provide design data applicable to typical utility installation. Emphasis has been placed on low to medium sulfur operation, with SO_2 removal efficiencies from 70 to 90 percent. Most test work has been in the operating mode where sorbent and fly ash have been recycled from the fabric filter back to the spray dryer, as is typical of many utility systems.

The pilot unit began operation in March 1982. This paper summarizes results from May 1982 through August 1983. Work during this period has included both once-through and recycle operation, inlet SO_2 concentrations of 400 to 2000 ppm, various modes of recycle, normal (270°F*) and low (210°F) inlet flue gas temperature, and varied makeup water compositions. In the following sections, the pilot unit is described, results are presented regarding the operability of the spray dryer and fabric filter, SO_2 removal results are presented, and the conclusions available as of August 1983 are summarized.

DESCRIPTION OF PILOT UNIT

The spray dryer unit at Arapahoe consists of a spray dryer vessel, one compartment of a fabric filter, a reagent preparation system, and reagent feed/recycle systems, as well as a comprehensive instrumentation system. Figure 1 is a simplified process flow diagram for the pilot unit as set up for rotary atomization.

The spray dryer vessel has a 10-foot diameter, 5.5 foot straight side and a 50° cone bottom. The vessel has interchangeable flue gas inlet vane rings (one for normal flow rates and one for low flow rates), can accommodate rotary or nozzle atomization, and can be rearranged for either bottom or side gas exit. An additional straight side section which doubles the volume of the dryer vessel can be installed to facilitate nozzle atomization tests. At normal flue gas rates (6500 to 9000 acfm dryer outlet gas rate), the vessel flue gas residence time without the added section is approximately 5 to 7 seconds. Thus, the vessel is quite versatile, allowing a number of potential configurations. This versatility was desired to provide the capability to include aspects of various vendors' designs in the test program.

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*British Engineering Units rather than SI units are used in this paper because of customary usage in the electric power industry. An appendix provides appropriate conversion factors.

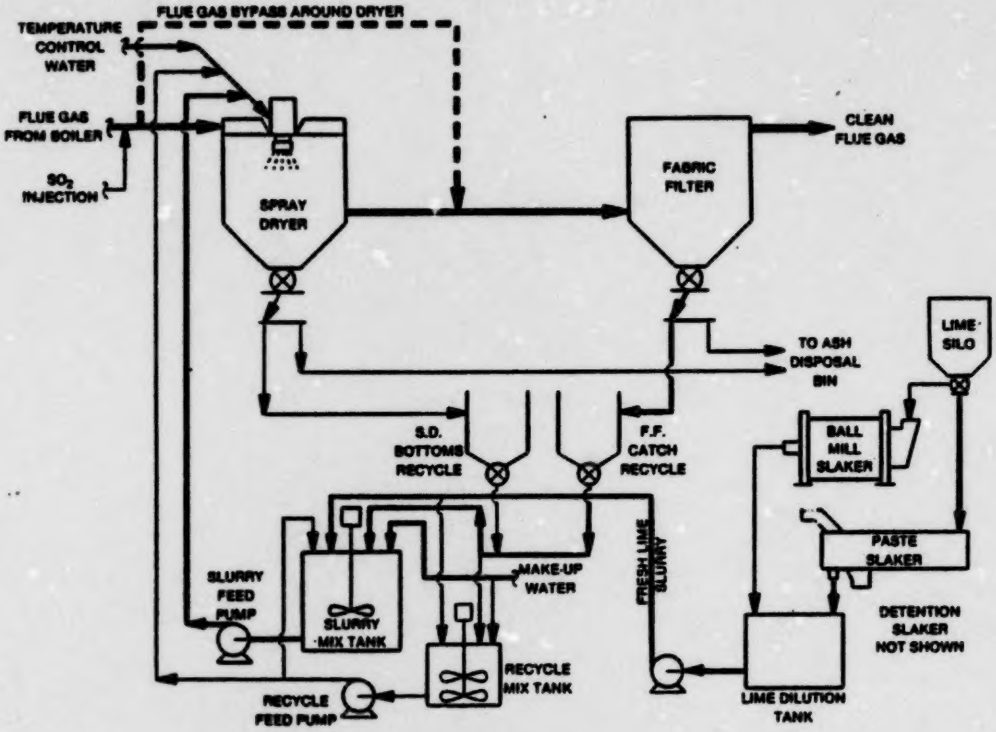

Figure 1. Simplified Process Flow Diagram for Spray Dryer Pilot Plant (Rotary Atomizer Configuration)

The fabric filter used for this spray dryer pilot plant is one compartment of a 10-MW equivalent four-compartment pilot unit which had previously been undergoing characterization for over a year. The fabric filter is also quite versatile in design, allowing a choice of bag diameters and bag lengths. The compartment is currently equipped with 36 utility-size (12 in. by 34 ft) glass fiber bags (tri-coat finish) in a 6 x 6 array. At the normal 6500 to 9000 acfm fabric filter module inlet flue gas rate, the air-to-cloth ratio varies from 1.7 to 2.3 ft/min. The unit has provisions for either reverse gas, shake/deflate, or combined cleaning. Each module, including the one used for the spray dryer pilot unit, is controlled by a microprocessor. The microprocessor enables full variation in cleaning cycle duration, cleaning intensity, null period durations, damper opening and closing rate, etc. So far in the program, the fabric filter has been operated on a 3-hour cycle, using reverse gas cleaning.

A recent modification to the pilot unit allows bypassing warm gas around the spray dryer to reheat the gas entering the fabric filter. The controls on this bypass line will allow either a constant reheat temperature or a constant bypass flow.

The reagent preparation area contains ball mill, paste-type, and detention slakers. Additionally, provisions are available to mix simulated cooling tower blowdown waters for use as slaking and/or dilution water. Lime slurry used by the spray dryer has primarily been slaked in the paste slaker, although some ball mill- and detention-slaked lime has been used.

Slurry is fed to the atomizer using progressing cavity pumps from one of two covered, well agitated tanks. Temperature control water is added at the atomizer. The system has the capability of recycling fly ash/spent sorbent from the spray dryer bottoms catch (during side flue gas exit operation) and/or from the fabric filter hopper cach. Solids from each location can be pneumatically transferred to separate recycle bins. Recycle material can either be slurried and fed from a separate recycle tank, or can be slurried in the main slurry feed tank along with fresh lime. The feed rates of recycle material and makeup water to either tank are ratioed to the rate at which fresh lime slurry is added to the slurry feed tank. A separate pump and controller are available to meter recycle material to the atomizer separate from the lime slurry if the two are not slurried together in the same tank.

The instrumentation on the pilot unit was discussed in detail in a previous paper, "EPRI Spray Dryer Pilot Plant Status and Results,"[1] given at the EPA/EPRI FGD Symposium in Hollywood, Florida, in May 1982.

OPERATION RESULTS

The EPRI pilot unit has provided insight into the operation of the equipment used in spray dryer/baghouse systems. Many improvements in pumps, tanks, agitators, etc., have been made since start-up to improve the operation of the pilot unit. However, most of these changes were necessary because of the small scale of this pilot plant and are not directly applicable to utility systems. However, observations about the operation and maintenance of the spray dryer and fabric filter are more generic in nature and may be applicable to most spray dryer/baghouse FGD systems. The spray dryer and fabric filter operation results are discussed in the following section.

SPRAY DRYER OPERATION

Operation of the system in the once-through mode, with lime reagent, for low inlet SO_2 levels (300 to 400 ppm) has been the most troublesome configuration tested to date. After an initial check-out of the system with soda ash reagent and a spinning disc atomizer wheel, the first lime slurry tests were conducted in the once-through mode. The results of soda ash testing and early once-through lime tests were reported in the previously referenced FGD symposium paper. The problems encountered after switching to lime slurry were multiple. First, the pilot spray dryer had been designed for a flue gas flow rate that resulted in a residence time of approximately 5 seconds. When trying to operate at a 20°F approach to adiabatic saturation at the dryer outlet, as many utility systems sold to date will, a continued buildup of wet solids on the dryer walls was encountered. This buildup caused several problems. First, wet solids from the walls would intermittently fall to the bottom of the dryer. The dryer is equipped for pneumatic conveying of solids from the bottom of the dryer, but these wet solids (10 percent moisture and greater) tended to plug the pneumatic lines. Once the pneumatic lines plugged, in spite of efforts to quickly clear the pluggage, wet solids would begin to bridge across the cone bottom on the dryer. This bridge of wet solids would continue to grow until the flue gas exit duct from the dryer was almost plugged. At this point, the dryer would have to be shut down, the rotary valve and pneumatic line removed from the bottom of the cone, and the vessel would have to be washed out. The problems of drop out of wet solids on the walls of the dryer were substantially reduced, however, when operating a lower flue gas flow rate (7 to 8 second residence time) or at a 30°F or higher approach to saturation at the dryer outlet.

The second major problem during the once-through lime tests was related to the atomizer rather than the vessel itself. For lime reagent tests, a nozzle insert-type wheel normally used for abrasive slurry service has been used. During the first several weeks of operation, the spray machine was taken off-line several times due to high vibration. Invariably, when the unit was removed for inspection, it was found that several of the twelve nozzles around the periphery of the wheel were plugged with lime solids. The pluggage problem was lessened but not solved by increasing the wheel speed from 12,000 to 15,000 rpm. It became apparent that the pluggage of nozzles with lime solids was caused by flue gas pumping through the wheel. The 9-inch diameter wheel with twelve 3/8-inch openings had a hydraulic capacity much greater than the nominal 2 gallons per minute of slurry normally fed at these flue gas rates. Because it had a greatly increased hydraulic capacity, it would educt hot flue gas across the top of the wheel to be pumped through the nozzles. Because of this flue gas pumping, lime and/or fly ash solids were deposited on the top side of the wheel, and lime solids tended to dry within the wheel, particularly within nozzle openings. An attempt was made to fit a seal between the spray machine and the wheel in order to restrict this pumping effect. However, the pumping continued, although perhaps at a reduced rate. As flue gas was pumped through this low clearance area, lime and/or fly ash deposited on the top of the wheel caused the seal on the bottom of the spray machine to rub the wheel. The resulting imbalances led to early failure of the lower bearing on the machine. An alternate approach, then, was to reduce the number of active nozzles in the wheel from twelve to six. The other six nozzles were replaced with solid inserts. Although the tendency for the nozzles to plug was reduced, on numerous occasions, three of the six nozzles would plug. If any two adjacent nozzles plugged, an imbalance would lead to high vibration measurements and cause the operators to shut the spray machine down. However, if every other of the six nozzles plugged, vibration problems were not encountered. this led to a further reduction in active nozzles to only three, where now nine of the original twelve nozzles were replaced with solid inserts. While this last modification solved the nozzle pluggage problem, it tended to aggrevate the previously discussed problem with wet product on the dryer walls. Empirical correlations in Masters[2] for predicting the droplet size produced by a vaned atomizer wheel suggest that for a nozzle wheel with a given nozzle diameter, increasing the number of nozzles around the periphery of the wheel will produce a finer droplet size. Based on this correlation, it would seem that in reducing the number of active nozzles in our atomizer wheel, we produce a coarser droplet size which is not as readily dried. This no doubt contributes to the buildup of wet product on the spray dryer chamber walls.

After once-through operation, lime reagent recycle tests were begun. In most recycle tests conducted thus far in the program, recycle material has been taken from the fabric filter hopper, and mixed with fresh lime and makeup water in the main slurry feed tank. This fresh lime/recycle slurry is normally fed to the atomizer by a single pump. Recycle ratios are expressed in this paper as pounds of recycle material per pound of hydrated lime ($Ca(OH)_2$) solids in the fresh lime feed. The lowest recycle ratios tested were around 2:1, but even at this level of recycle an immediate improvement in spray dryer operation resulted. With recycle, the number of active nozzles in the atomizer wheel could be increased back to six from the three required for once-through operation. This increase in active nozzles should result in somewhat finer atomization, further promoting improved drying.

Upon the implementation of recycle, problems with buildup of solids within the atomizer wheel, buildup of wet solids on the spray dryer chamber walls, and plugging of the pneumatic transfer line from the spray dryer bottom have also been virtually eliminated. The dryer can be successfully operated over a long term at a 20°F approach, at a flue gas flow rate resulting in a 7-second dryer residence time. At recycle ratios near 12:1, resulting in a solids level at the atomizer wheel of 35 to 45 percent, the dryer has been successfully operated for days at a time at a 20°F approach at residence times as low as 5 seconds.

It appears that much of the improvement in spray dryer operability in recycle operation is related to having an increased weight percent solids level in the slurry fed to the atomizer. This effect has become apparent in recent once-through lime tests conducted at 1000 ppm inlet SO_2 levels. At 1000 ppm inlet levels, roughly three times the amount of lime is fed compared to low sulfur operation, at equivalent reagent ratios. This solids level, then, is equivalent to the solids level at the same reagent ratio for 350 ppm inlet SO_2 and 2:1 recycle ratio operation.

In the baseline 2:1 recycle tests, the weight percent solids in the slurry atomized ranged from 6 to 16 percent. Similarly for the 1000 ppm inlet SO_2 once-through lime tests, the weight percent solids level to the wheel varied from 12 to 16 percent. In these once-through tests at a nominal 7-second dryer residence time, problems of wheel pluggage with the six nozzle atomizer wheel, and pluggage of the drying chamber with wet solids were not encountered. These successful once-through lime tests at a 1000 ppm inlet SO_2

level seem to indicate that at least part of the improved spray dryer operability during recycle operation comes from the increased solids level in the slurry atomized rather than from the presence of the recycled material per se.

Further evidence of this weight percent solids effect on operability is seen for the high recycle rate tests. For the nominal 12:1 recycle tests the percent solids to the wheel ranged from 31 to 37 percent. Operability of the spray dryer was improved over that even for nominal 6:1 recycle ratios. At an inlet flue gas flow rate of 5800 scfm (5.0-second residence time in dryer) when recycling to a nominal 35 weight percent solids level to the wheel, no tendency to drop out wet solids in the dryer or plugging of pneumatic transfer lines was noted. However, some tendencies for wet product formation had been seen when operating at this flue gas rate at lower recycle ratios (hence lower weight percent solids to the wheel).

The effects of operating condition on the weight percent solids in the slurry fed to the atomizer are summarized in Table 1. The table shows that once-through operation at low inlet SO_2 concentration results in the lowest solids content in the slurry fed to the atomizer. It was at these conditions that problems with wet spray dryer products persisted. Other operating modes resulting in higher solids content in the slurry feed (recycle, and/or higher inlet SO_2 concentration) greatly improve the spray dryer operation.

TABLE 1. SUMMARY OF WEIGHT PERCENT SOLIDS IN SLURRY FED TO ATOMIZER WHEEL AT VARIOUS OPERATING CONDITIONS

Operating Mode	Nominal Inlet SO_2 (ppm)	Weight Percent Solids Atomizer Wheel	Vessel Residence Time (sec)	Nature of S.D. Product
Once-through	350	3 to 7	7	Wet
Recycle (2:1)	350	6 to 16	7	Dry
Recycle (6:1 to 8:1)	350	17 to 24	7	Dry
Recycle (12:1)	350	31 to 37	5	Dry
Once-through	1000	12 to 16	7	Dry
Recycle (2:1)	1000	20 to 34	7	Dry
Recycle (2.5:1)*	1000	25*	5	Dry

*One test.

Theoretical considerations presented by Masters[2] would tend to support these observations of improved operability at higher feed solids contents. Masters describes the first two phases of drying of droplets in a spray dryer as the "constant rate" period (the first phase of drying) and the "first falling rate" period (the second phase of drying). In the first phase, or constant rate period, the surface of the droplet behaves as though no solids are contained, and water evaporates freely from the surface. In the first falling rate period (second phase of drying), solids protrude from the surface of the droplet, and the rate of evaporation of moisture from the droplet is slowed by the rate at which water diffuses through the network of solid particles.

It is important that droplets atomized are beyond the constant rate, or first drying period before they come into contact with vessel walls. The length of the first drying period is labeled as the critical drying time. In general, problems with wet product drop out in the dryer can be avoided by decreasing this critical drying time for the droplets atomized. Empirical correlations presented in Masters allow one to predict the impact of variables such as solids content on critical drying time. The results of example calculations are summarized in Table 2. These example results show that for a 50-micron droplet, the critical drying time is reduced by one-half if the solids content is raised from 5 to 30 percent. Calculations for 100-micron droplets show that critical drying times are 3 to 4 times those for the 50-micron droplets, but that the higher solids level again nearly reduces the critical drying time by one half. Of course a higher solids content in the slurry feed to an atomizer will also impact the atomized slurry particle size distribution to a minor degree, but it is clear from these example calculations that the net effect of increased solids content in the slurry feed is reduced critical drying time, hence potentially improved dryer operation. Also in Table 2 are falling rate drying times and total drying times for each of the four cases. The falling rate drying times and total drying times actually increase at higher weight percent solids levels. This increase may account for some of the improved SO_2 removal performance seen in recycle operation. Please note that the values presented in Table 2 were calculated for example only, and do not represent an attempt to accurately predict the critical and/or total drying times in the Arapahoe spray dryer.

TABLE 2. EXAMPLE DRYING TIMES FOR ATOMIZED DROPLETS

	Case 1	Case 2	Case 3	Case 4
Droplet Size, μ	50	50	100	100
Weight Percent Solids	5	30	5	30
Critical Drying Time, sec	0.4	0.2	1.4	0.8
Falling Rate Drying Time, sec	0.1	0.8	0.4	2.0
Total Drying Time, sec	0.5	1.0	1.8	2.8

BAGHOUSE OPERATION

Throughout over 6000 hours of spray dryer operation, no bag fabric problems have been observed. This result is significant, because the operating time includes both sodium carbonate and lime reagent, once-through and recycle operation, low sulfur to medium sulfur inlet SO_2 levels, a great number of start-ups and shutdowns, and several spray dryer upset conditions. Spray dryer upsets which one would have expected to have a detrimental impact on bag performance did not seem to affect the bags at all. In one incident, a new operator flushed a slurry feed line into the spray dryer through the atomizer, driving the dryer outlet temperature to the adiabatic saturation temperature. In spite of such abuse, no bag failures have occurred, outlet opacity remains quite low, and tube sheet pressure drop values have been quite acceptable.

Although there have been no bag fabric problems due to operation downstream of the spray dryer per se, there has been a problem with corrosion on the baghouse walls. The problem did not become apparent until after the spray dryer had been in operation for approximately 8 months, in November 1982. During an outage, the baghouse module was opened and large flakes of rust scale were noted to be forming along the inside walls. The corrosion was attributed to moisture condensation on compartment walls. Several factors may be contributing to this condensation and corrosion problem. First, the compartment is part of a pilot unit, so it goes through numerous start-ups and shutdowns, passing through the dewpoint of the flue gas each time. In addition, since the compartment shares no common walls with other compartments, all four walls of the compartment are exposed to ambient temperatures. Also, the baghouse was originally designed to clean hot gas (250°F to 300°F) rather than gas which is approaching adiabatic saturation. Consequently, the baghouse walls are insulated with only a 2-inch thickness of fiberglass batts. At the spray dryer outlet, the flue gas is generally 17 to 22°F from adiabatic saturation. This corresponds to roughly 20 to 30°F from the dewpoint. As the flue gas is cooled in the outlet ductwork and in the baghouse, the bulk gas temperature may come within 10 to 15° of its dewpoint. Any signficant cooling of the baghouse walls, then, may put the skin temperature at or below the dewpoint of the flue gas. This effect would particularly be noticeable in very cold weather, approaching 0°F, where the driving force for cooling the walls would exceed 100°F. The explanation that the corrosion is caused by localized cold spots is supported by the observation that corrosion is worst near an entry door to the compartment, where not only is the wall insulation sparse, but where cold air can leak into the compartment.

Sulfuric acid condensation is not thought to directly contribute to the corrosion problem, because no measurable SO_3 content has been found in the flue gas downstream of the filter bags. However, SO_2 present in the outlet flue gas would be sorbed into any condensation which forms. Subsequent liquid phase oxidation reactions would tend to form sulfuric acid in the condensed moisture. Sulfuric acid-laden moisture then would not readily evaporate because of the azeotropic nature of the sulfuric acid-water mixtures.

A later outage, after approximately 10 months of operation, showed continued corrosion. Additionally, during this outage the top side of the tube sheet was found to be covered to a depth of several inches with wet ash and flakes of corrosion. The moisture on the tube sheet was attributed to the severe spray dryer upsets mentioned previously, where an inexperienced operator repeatedly overloaded the atomizer with water while flushing a line. The ash on the tube sheet had been observed during a previous outage. It is not clear when this dry ash was deposited. The top of the tube sheet was cleaned and the unit operators were reminded of proper line flushing procedures which call for blocking out flow to the atomizer.

During this outage, thermocouples were installed on two wall surfaces within the fabric filter compartment and continuously monitored. In colder weather, it was observed that the temperatures were generally several degrees below the bulk gas temperature. In April 1983, the ambient temperature overnight in Denver dropped to 9°F. One of the two wall temperatures dropped to 96°F for over an hour. This is several degrees below the dewpoint of the bulk gas at the fabric filter outlet. The pilot unit was shut down and the compartment was opened. As expected, moisture was found on compartment walls and on top of the tube sheet. More startling was the discovery of three bags collapsed onto the tube sheet. The bags had fallen due to corrosion failure of the mild steel bag caps. These bag cap failures were not thought to be the result of this one low temperature excursion, but a cumulative effect of 13 months of operation downstream of the spray dryer and a particularly cold winter.

The bags were all removed and replaced, and the removed bags were examined. A map of bag cap corrosion was prepared, and is presented in Figure 2. In general, the caps showing the worst corrosion were around the edges. Of the three failed bags, two were in corners and one was near the compartment doors. Top and bottom portions from twenty of the bags were submitted to Albany International for routine performance evaluation. No failures or areas of potential failure were observed on any of the submitted samples except a top sample from one of the fallen bags. This top sample contained two small holes in an area which was heavily stained with rust. The damage likely occurred when the bag fell.

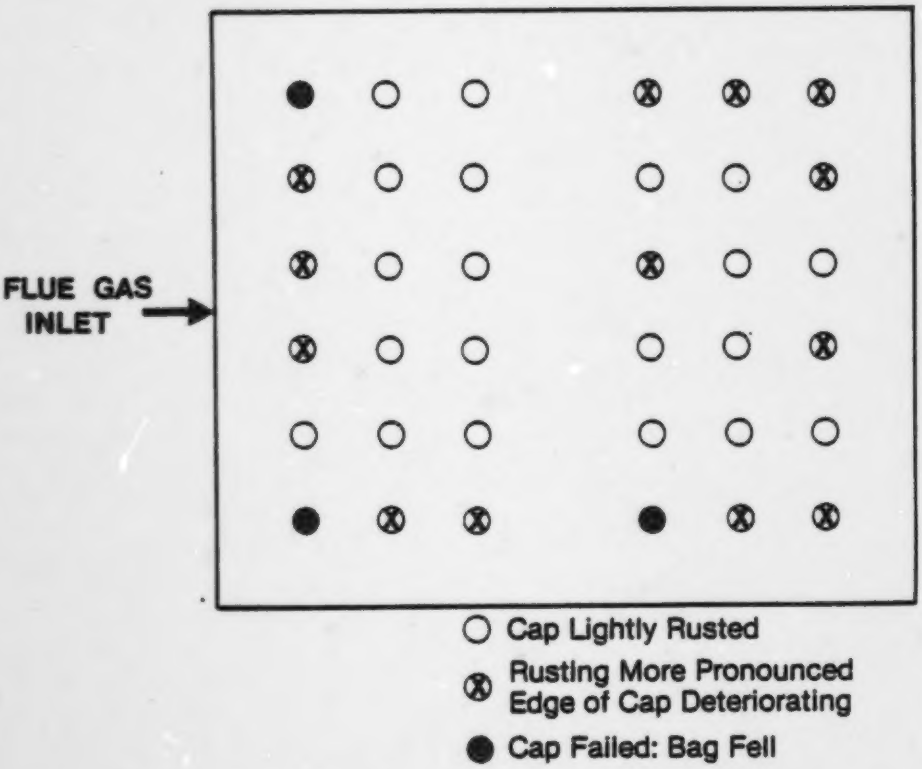

Figure 2. Map of Bag Cap Corrosion/Failure in Compartment A

During this outage, the compartment was instrumented with a total of 10 thermocouples. These thermocouples monitor various wall and bag cap temperatures. A map of the thermocouple locations is shown in Figure 3. The east and south facing walls of the compartment are exterior walls of the fabric

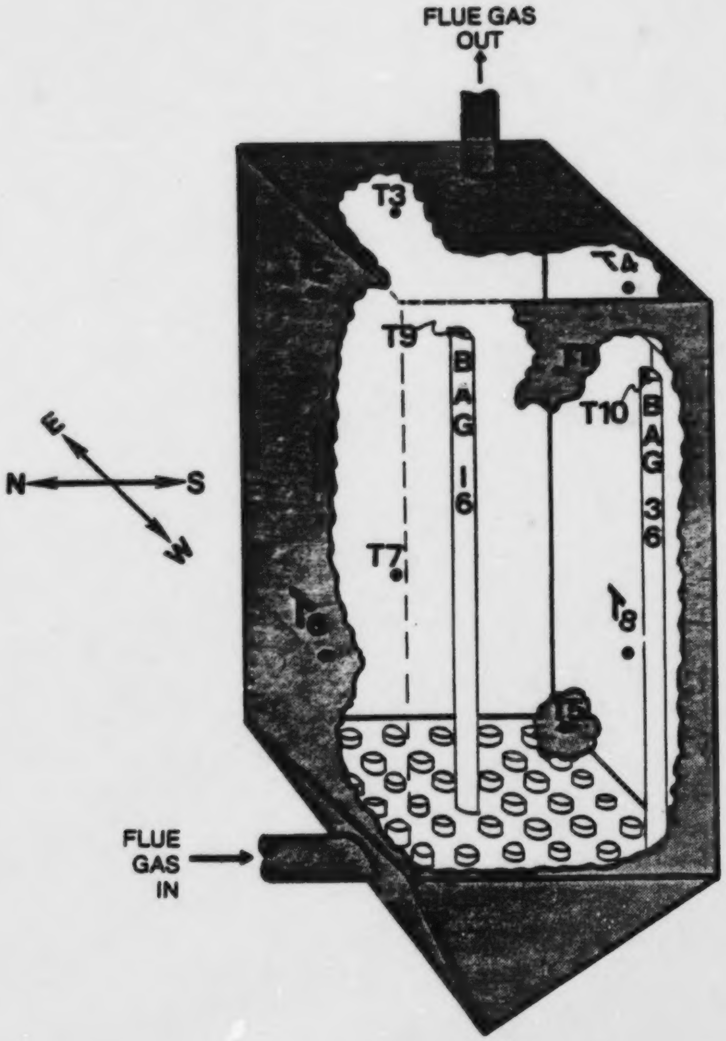

Figure 3. Cutaway View of Thermocouple Locations in Baghouse Compartment

filter enclosure, while the west and north walls face the inside of the enclosure. Table 3 summarizes selected temperature measurements since the pilot unit was put back in service. Case 3 in the table includes the two wall measurements previously recorded for a 9°F ambient temperature. Note that in Case 1 and Case 2, no temperatures are actually below the bulk gas dewpoint, but for Case 2 (36°F ambient) one of the wall temperatures is approaching that value. Also note that the T_5 temperature which reached 96°F in Case 3 is not even the coolest temperature in the other two cases. Monitoring of all ten locations during the upcoming winter is expected to identify a number of excursions of wall and cap temperatures below the bulk gas dewpoint.

TABLE 3. AMBIENT TEMPERATURE EFFECTS ON BAGHOUSE TEMPERATURES

	Temperature (°F)		
	Case 1	Case 2	Case 3
Ambient	82	36	9
S.D. Outlet	138	135	137
Baghouse Outlet	132	124	125
Approximate Dewpoint (F.F. Outlet)	108	104	104
T_1 (Top Interior Wall)	117	107	-
T_5 (Bottom Interior Wall)	125	113	96
T_8 (Bottom Exterior Wall)	125	117	109
T_9 (Bag 16-Cap)	136	130	-
T_{10} (Bag 36-Cap)	133	124	-

These results certainly indicate that full-scale baghouses downstream of spray dryers will have to be designed with ample insulation on exterior walls, and should be insulated between compartments to avoid cold walls when adjacent compartments are off-line. Also, compartment doors must be well sealed, and penetrations of supports through the insulation and welding directly to compartment walls should be avoided.

Replacement of the bags in the compartment downstream of the spray dryer has allowed a number of "before" and "after" comparisons. Since this compartment is one of four in the same baghouse, a control compartment which treats the same flue gas without the spray dryer upstream is also available for comparison.

The first comparison (Figure 4) shows the pressure drop across Compartment A (downstream of the spray dryer) compared to that for Compartment C (control, fly ash only) during the month of January 1983. In this plot, the original bags were still in service downstream of the spray dryer. The lower pressure drop across Compartment A is expected considering Compartment A was operated at a lower filtering air-to-cloth ratio of 1.7 ft/minute compared to 2.0 ft/minute in Compartment C. This result is somewhat surprising though, considering that Compartment A is removing 2 to 3 times the quantity of particulate matter downstream of the spray dryer than Compartment C collects from the untreated flue gas.

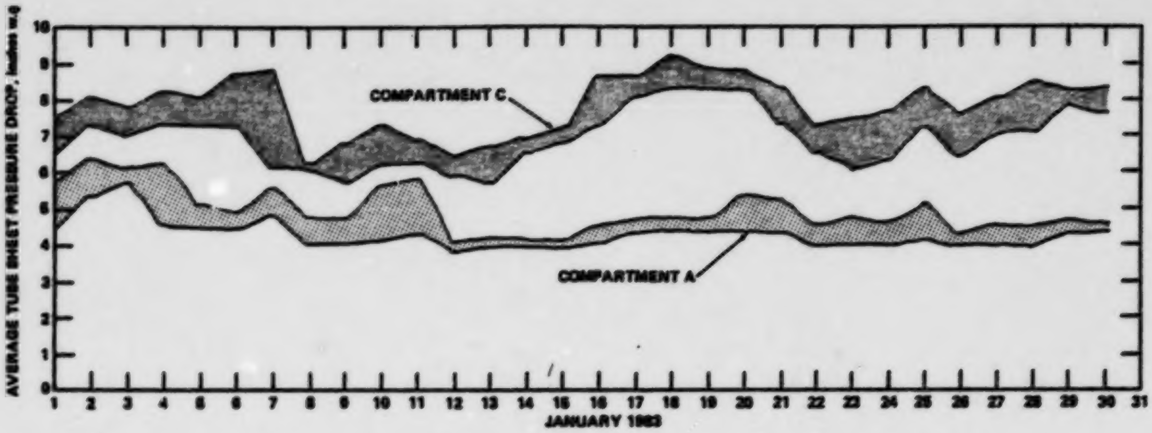

Figure 4. Average Tube Sheet Pressure Drop History - Compartments A and C - January 1983

The bags in Compartment A were originally filtering fly ash only for approximately 5 to 6 months prior to start-up of the spray dryer (July 1981 - January 1982), thus establishing a residual fly ash cake on the bags. When new bags were placed in service after the corrosion problem in April 1983, the spray dryer was started up immediately; no conditioning on fly ash-only was conducted. In Figure 5 a pressure drop comparison of Compartments A and C, 4 months after the new bags were installed in Compartment A, for July 1983, shows a substantially reduced pressure drop across Compartment A relative to that of Compartment C. When the difference in A/C ratios is accounted for, the Compartment A Δp is still significantly lower than that for Compartment C. The difference is explained by comparing the period of operation between the two compartments. The Compartment C bags have been in operation for over two years and have reached a steady-state pressure drop and residual dust cake weight (50-60 lbs). The Compartment A bags, after operating only 4 months, have a very low pressure drop and only about 15 lb of residual dust cake. Whether the new Compartment A bags will eventually reach a Δp and residual cake weight similar to Compartment C remains to be seen. However, short-term operation without fly ash conditioning has not proven detrimental. In fact, the Δp across Compartment A has risen very slowly and does not appear to be substantially increasing with time. This effect is shown in Figure 6.

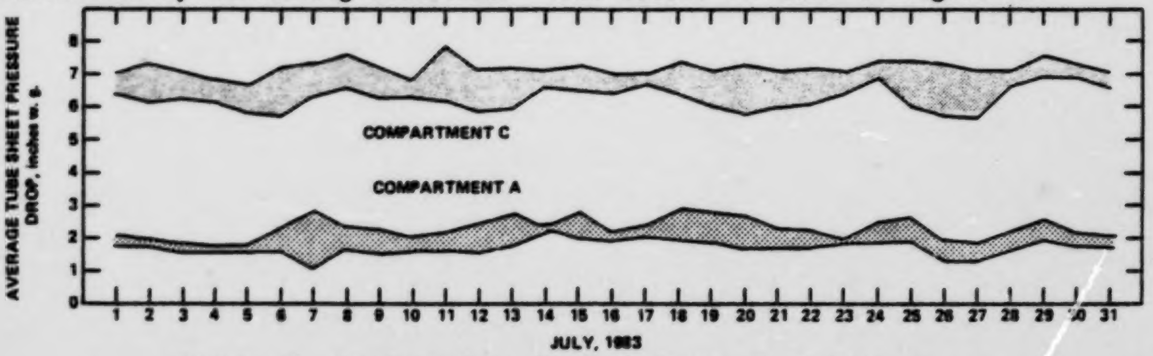

Figure 5. Average Tube Sheet Pressure Drop History - Compartments A and C - July 1983

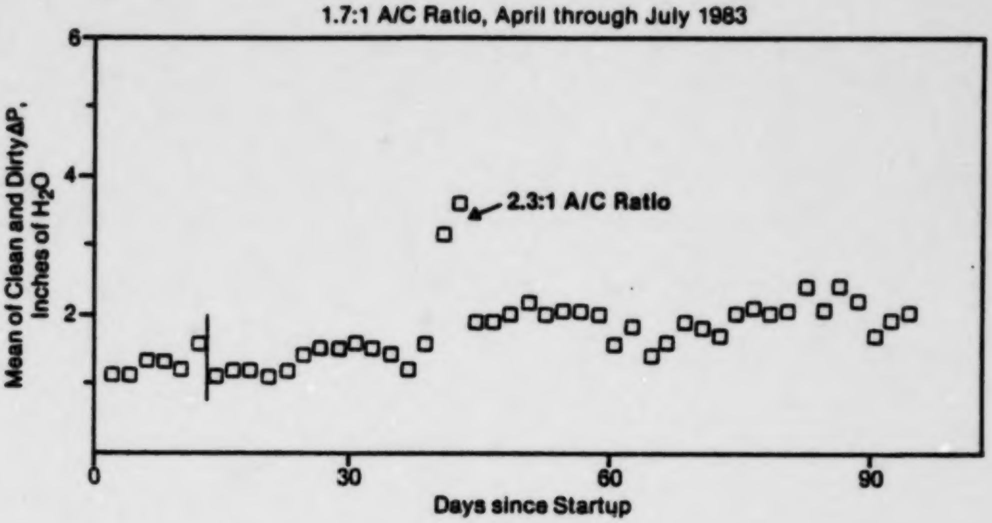

Figure 6. Pressure Drop Profile of a Recent Start-Up of Compartment A Downstream of Spray Dryer

Several important findings have resulted from the Arapahoe pilot baghouse operation downstream of the spray dryer. These include:

1) Operation downstream of a spray dryer does not appear to have an adverse affect on the bag fabric, even when subjected to occasional upset conditions at or very near adiabatic saturation;

2) A baghouse downstream of a spray dryer operated at close approach to adiabatic saturation must be well insulated, with close attention paid to installation details in order to avoid cold spots which lead to localized corrosion; and

3) A baghouse downstream of a spray dryer may operate at substantially lower pressure drop than the same baghouse treating normal fly ash-laden flue gas, particularly if the bags are not conditioned by any lengthy period of fly ash-only operation.

SO_2 REMOVAL RESULTS

The SO_2 removal results presented here concentrate on the recycle lime reagent operating mode. This mode is typical of the majority of utility spray dryer/fabric filter dry FGD systems sold to date, and represents the most operable configuration for this pilot unit. The recycle data presented here all represent recycling solids from the fabric filter.

In most of the data plots presented in this section, SO_2 removals across the overall system, across the spray dryer alone, and across the fabric filter are plotted for each test. SO_2 removal across the fabric filter in these plots is calculated by the following equation:

$$\text{SO}_2\text{ Removal in Fabric Filter} = \frac{(\text{SO}_2\text{ Spray dryer outlet}) - (\text{SO}_2\text{ Fabric filter outlet})}{(\text{SO}_2\text{ Spray dryer inlet})} \times 100\%$$

By calculating SO_2 removal across the fabric filter in this manner, the SO_2 removal across the spray dryer and across the fabric filter can be directly summed to get the overall system removal. The SO_2 removal values for the fabric filter presented here represent time averaged values, which have a number of 3-hour cleaning cycles included in the averaging time. This time averaged value is thought to coincide well with the instantaneous values one would encounter across a multicompartment baghouse, where perhaps ten compartments would be evenly distributed with respect to time throughout their cleaning cycles.

Each data point for the recycle tests represents the mean value for approximately 24 to 48 hours of steady-state operation at those conditions. Steady-state operation is normally achieved in approximately 24 hours. For the system to be at steady state, the bags in Compartment A must have been cleaned, the Compartment A hopper emptied, the recycle material holding bin refilled, and the combined lime/recycle slurry feed tank refilled at least two to three times each. When steady state is achieved, the chemical composition of the slurry fed to the atomizer and of the material collected on the bags in Compartment A no longer change significantly from batch to batch.

Baseline SO_2 removal results for recycle operation at normal Arapahoe inlet SO_2 concentrations around 350 to 400 ppm are presented in Figure 7. These results represent an approach to adiabatic saturation at the dryer outlet within 20°F, and a recycle ratio of 2:1. The recycle ratio was defined earlier as:

$$\text{Recycle ratio} = \frac{\text{lb of material recycle (dry basis)}}{\text{lb of fresh Ca(OH)}_2\text{ makeup (dry basis)}}$$

The results in Figure 7 show that at low reagent ratio values (below 0.8) fabric filter SO_2 removal contributes little to overall SO_2 removal, because lime reagent utilization in the spray dryer approaches 100 percent. At higher reagent ratio values (1.3 and greater) fabric filter SO_2 removal contributes greatly to system SO_2 removal, because SO_2 removal in the spray dryer does not increase with any significance as the reagent ratio is increased. In this paper, unless otherwise noted, reagent ratio is defined as:

$$\text{Reagent Ratio} = \frac{\text{lb-mol fresh lime fed to spray dryer}}{\text{lb-mol SO}_2\text{ in inlet flue gas to spray dryer}}$$

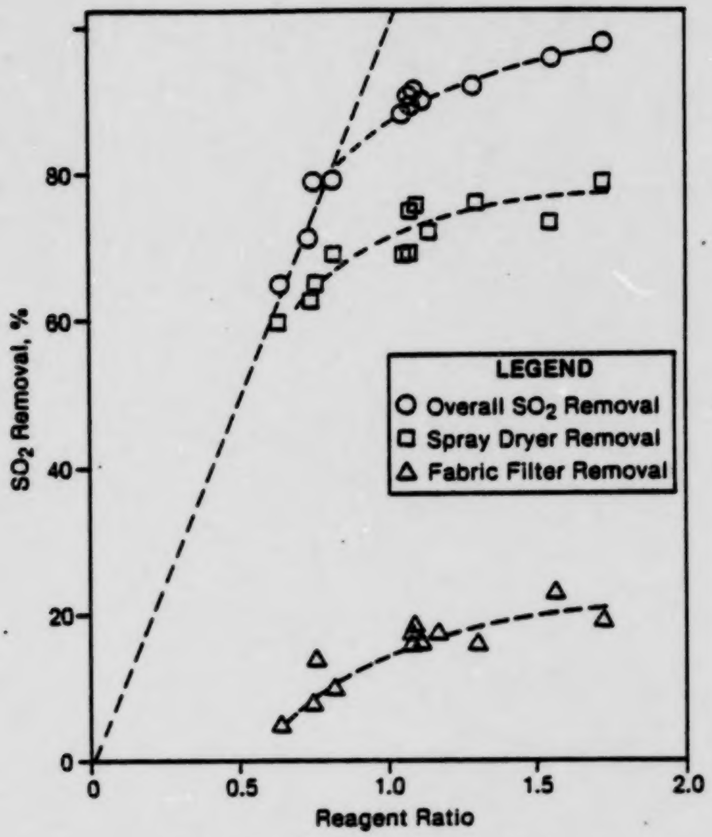

Figure 7. Baseline SO_2 Removal for 400 ppm Inlet SO_2, 20° Approach, 2:1 Recycle Ratio

Once baseline performance curves for moderate recycle rates were determined a series of tests at higher recycle rates was conducted. These results are plotted in Figure 8. Recycle at ratios of approximately 6:1 to 8:1 did not appear to demonstrate any marked improvement in SO_2 removal over that for recycle ratios around 2:1. Recycle ratios of approximately 12:1 did show a marked improvement, primarily in SO_2 removal by the spray dryer. These highest recycle ratios correspond to solid contents of 30 percent or better in the slurry fed to the atomizer. The fabric filter contribution to overall removal does not appear to be affected by the increased recycle ratio. The upper curves in Figure 8 represent a visual fit of the SO_2 removal data across the spray dryer during 12:1 recycle operation. These upper curves show virtually complete lime utilization up to a reagent ratio of 0.9.

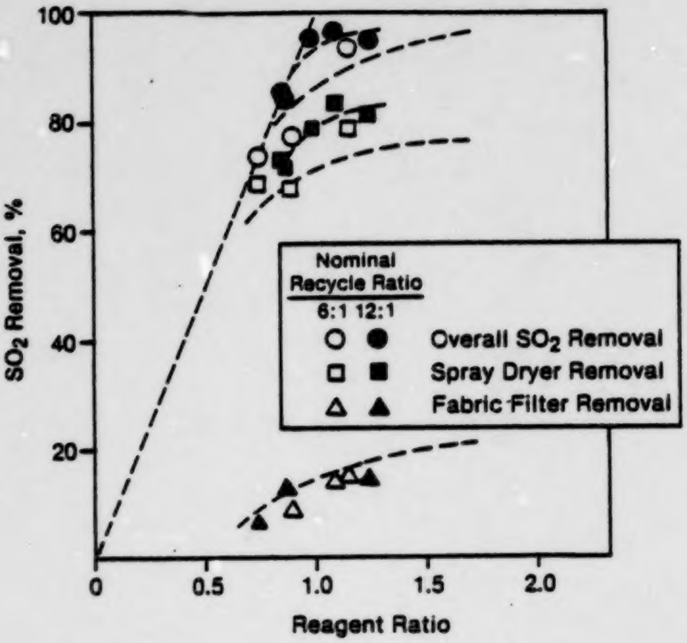

Figure 8. SO_2 Removal Results for Recycle Tests - 20°F Approach Temperature

Mention was made earlier in this paper that operation of the spray dryer/fabric filter FGD system is improved by recycle versus once-through lime operation. Both spray dryer operability and SO_2 removal improvements have been noted. The results of once-through lime tests are compared to the 2:1 recycle, 20°F approach, 350 to 400 ppm inlet SO_2 level baseline test results in Figure 9. Because of the repeated operability problems during once-through operation, many once-through SO_2 removal test results have been discarded. The results in Figure 9 represent a number of successful short-term once-through tests. The open symbols represent once-through results at a 20°F approach to saturation at normal Arapahoe inlet SO_2 levels. The dashed curves are the visual best fit of the 2:1 recycle, 20°F approach baseline results. The figure shows that at low reagent ratio (below 1.0), the difference in SO_2 removal performance between once-through and recycle operation is difficult to quantify because of scatter in the results of only two successful short-term tests. The results at higher reagent ratios (nominally 1.5) show that once-through operation gives approximately 10 to 15 percent lower SO_2 removal than for recycle operation. This difference is primarily due to a decrease in spray dryer rather than fabric filter SO_2 removal performance during once-through operation. It may have been that more successful once-through tests were completed at higher reagent ratios because of the beneficial impacts on spray dryer operability of higher weight percent solids in the lime feed for these tests.

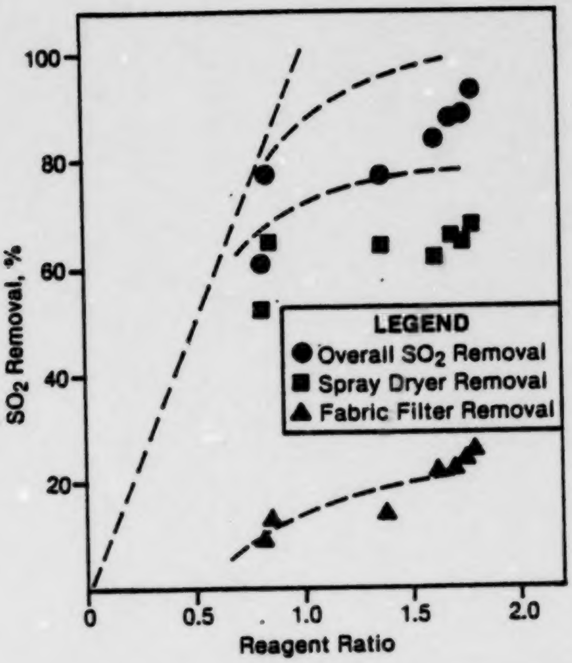

Figure 9. Comparison of Once-Through SO_2 Removal Points with Best Fit Curves from Recycle Results for 400 ppm Inlet SO_2 Levels

With the performance of the spray dryer/fabric filter system well characterized for a nominal 350 to 400 ppm inlet SO_2 level, an extensive series of tests at a 1000 ppm inlet SO_2 level was begun. Testing at this inlet SO_2 level was of interest because it simulates performance of the system on medium sulfur coals (1 to 2 percent sulfur). Also, it was felt that operation at higher inlet SO_2 levels would make subtle variable effects on SO_2 removal performance more pronounced.

Figure 10 summarizes the results of a number of tests conducted at a 1000 ppm inlet SO_2 level, with a 2:1 to 4:1 recycle ratio and a 20°F approach to adiabatic saturation. Also plotted is a "best fit" line for the 400 ppm results at similar conditions. The majority of the test results are in the 80 to 90 percent SO_2 removal range, which is what current utility boiler New Source Performance Standards would require for most medium sulfur content coals. The figure shows that at moderate overall SO_2 removal levels (70 to 85 percent), somewhat lower SO_2 removals result compared to what would result at an equivalent reagent ratio in the 400 ppm tests. At around 90 percent SO_2 removal, the overall SO_2 removal curves tend to intersect, with a reagent ratio of approximately 1.3 required to achieve 90 percent removal in either case. The shape of the overall performance curves in each case appears to be set by spray dryer removal performance. The fabric filter contribution data appear to plot on one line regardless of inlet SO_2 concentration. The intersection of the two spray dryer SO_2 removal curves appears to be the result of approaching an observed maximum for SO_2 removal across this particular spray dryer. It is not obvious whether this observed maximum, at around 78 to 80 percent, is the result of limitations on the effectiveness of flue gas SO_2 and slurry contact in this particular vessel, or is perhaps dependent on some other parameter, such as atomized slurry particle size distribution.

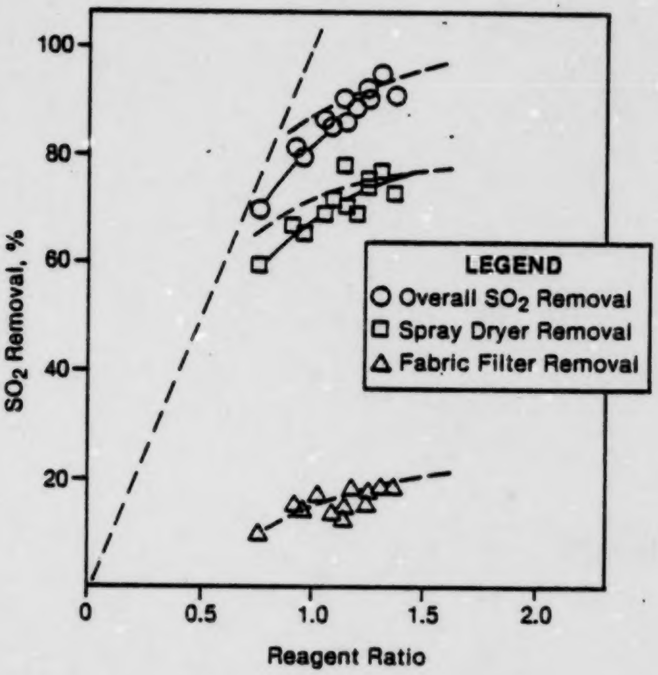

Figure 10. Comparison of 1000 ppm Inlet SO_2 Recycle Results with Best Fit of 400 ppm Inlet SO_2 2:1 Recycle Test Results

Once-through tests at 1000 ppm inlet SO_2 levels were also conducted. These results are compared to the 1000 ppm recycle results in Figure 11. Contrary to the 400 ppm inlet SO_2 test results, the difference between once-through and recycle tests at 1000 ppm inlet SO_2 levels is less dramatic. For a given reagent ratio, SO_2 removal in once-through operation is generally within 10 percent of removal in recycle operation. A reason for this observation may be the greatly increased weight percent solids to the wheel in once-through operation when the inlet SO_2 levels are increased by a factor of 2-1/2 to 3. For example, once-through operation at 400 ppm or lower inlet SO_2 results in a lime weight percent of only 3 to 7 percent at the atomizer. Similar reagent ratios for once-through operation at 1000 ppm inlet SO_2 levels result in weight percent solids levels of 10 to 16 percent. These solids levels are the same as solids levels at 400 ppm inlet SO_2 for 2:1 recycle operation.

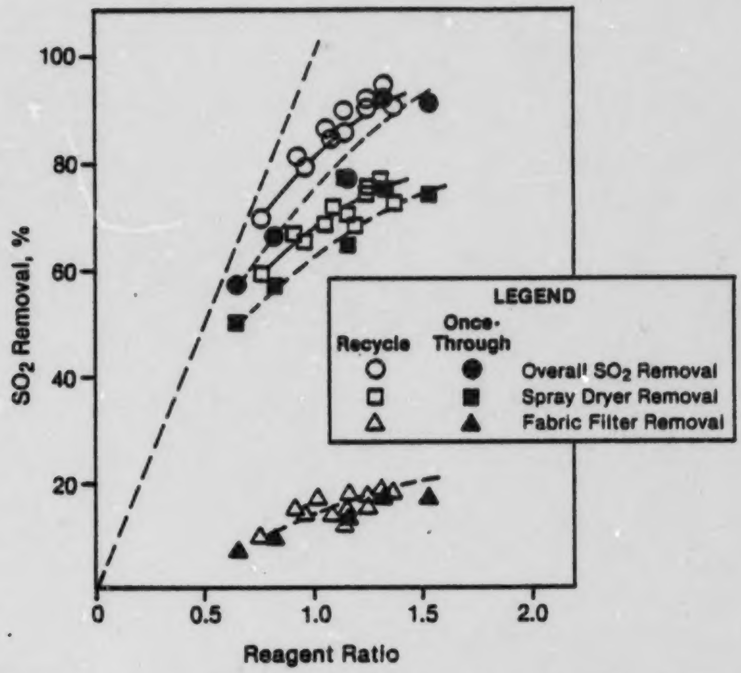

Figure 11. Comparison of Recycle Versus Once-Through Operation Results for 1000 ppm Inlet SO_2

Improved once-through SO_2 removal performance at higher inlet SO_2 levels is further seen in Figure 12, where once-through SO_2 removal results at both inlet SO_2 levels are plotted. In this figure, it can be seen that except for one 400 ppm test at low reagent ratio, all of the 1000 ppm test results for overall and spray dryer SO_2 removal plot on a curve above that for the 400 ppm test results. The one 400 ppm inlet SO_2 test result at close to 100 percent utilization at an overall removal level of 75 to 80 percent is suspected to be invalid. Current heat balance analyses of all test results should serve to either substantiate the results of this test or indicate that the point should be repeated.

A number of tests were conducted at low inlet temperatures of 210°F to 220°F. At Arapahoe barometric pressure conditions, this results in a spray down temperature of only 100°F to 110°F. For normal inlet temperatures of 260°F to 280°F, the spray down temperature is on the order of 145°F to 165°F. The 400 ppm inlet SO_2 tests in recycle operation are of particular interest at reduced inlet temperature. It is for low sulfur content coal, resulting in FGD inlet SO_2 levels of 400 ppm and less, that low inlet temperatures are likely to be encountered. Four tests were conducted at low inlet temperatures in the 400 ppm, recycle mode. These test results are plotted with the 400 ppm, 2:1 recycle, normal inlet temperature baseline results in Figure 13. The four low inlet temperature tests are at a wide range of reagent ratios and show a good bit of scatter. However, they show that SO_2 removal performance is not substantially reduced at low inlet temperatures. In fact, one point demonstrates overall SO_2 removal in excess of 90 percent at a reagent ratio between 1.2 and 1.3.

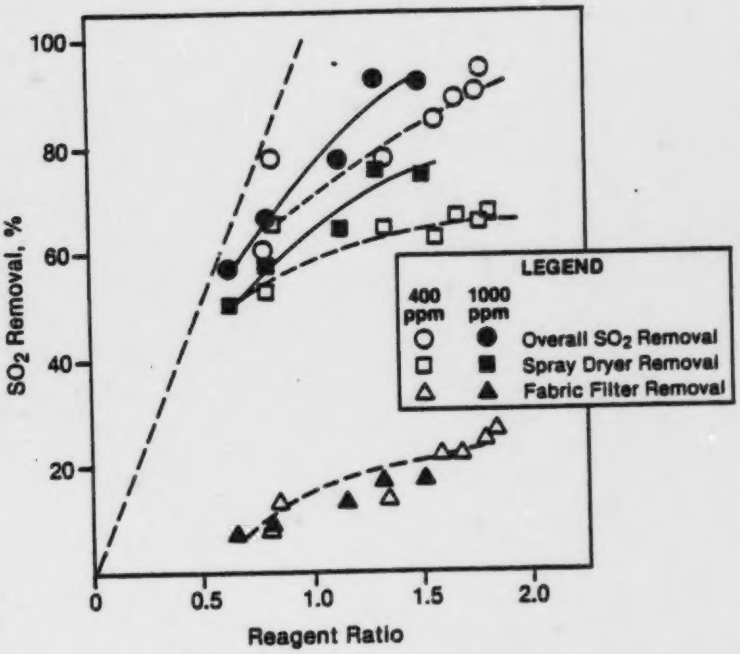

Figure 12. Comparison of Once-Through Lime Removal Results for 400 ppm and 1000 ppm Inlet SO_2 Levels

The following are two possible explanations for these results. One is that for low inlet SO_2 values, a 20°F approach to adiabatic saturation, and recycle operation, a reactive lime will be effectively utilized for SO_2 removal regardless of the spray down temperature. Another consideration is that other things being equal (inlet SO_2 concentration, reagent ratio, recycle ratio), lowered inlet temperature will increase the weight percent solids in the slurry fed to the atomizer. The results in Figure 13 may indicate that percent solids in the feed slurry has as important a role in SO_2 removal as does spray down temperature.

A consequence of the recent replacement of bags in the fabric filter compartment downstream of the spray dryer is that a comparison can be made of SO_2 removal performance of high residual ash, higher pressure drop old bags versus low residual ash, lower pressure drop new bags. The question to be answered by these test results was whether or not the nature of the permanent ash buildup impacts the effectiveness of the bags as SO_2 removal devices. It was speculated that the new bags, with less permanent ash buildup, would tend to have flue gas flow and fresh ash buildup more evenly distributed across the bag surface. This might improve the SO_2 removal performance of the fabric

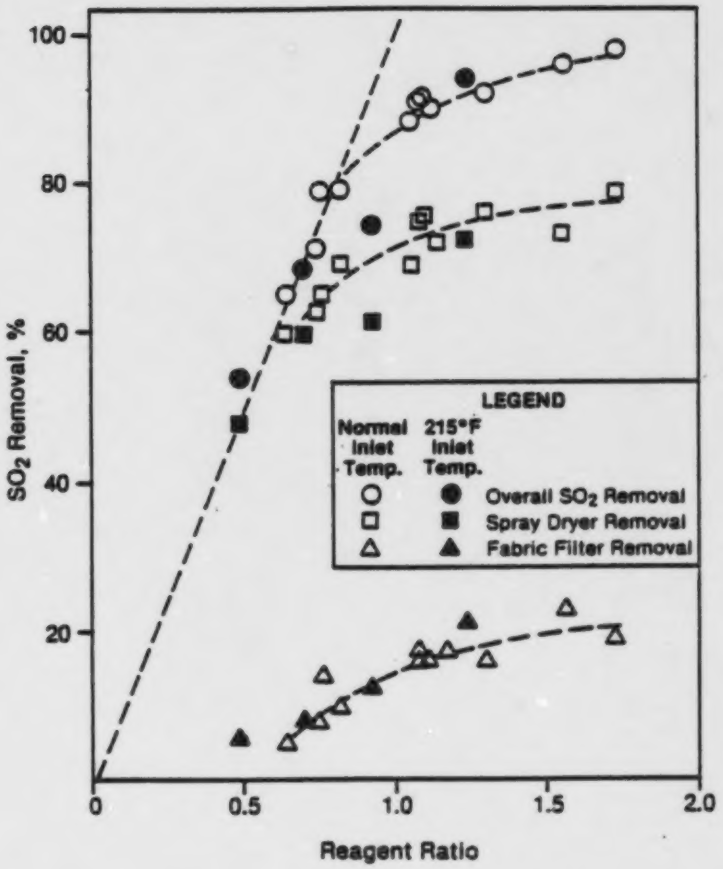

Figure 13. Effects of Reduced Inlet Temperature on SO_2 Removal for 400 ppm Inlet SO_2, Recycle Operation

filter after the new bags were installed. However, the results in Figure 14 do not support this speculation. These results compare 1000 ppm inlet SO_2, recycle test results with the old bags to similar tests after the new bags were put in service. The open symbols indicate the previous results, while the darkened symbols represent results with the new bags. The tests were conducted over a time period from immediately upon startup of the new bags to several months afterward. No dependence of overall or fabric filter SO_2 removal on old versus new bags or on operating hours on the new bags is apparent in these results.

Crucial to the use of spray drying technology for many western plants is the ability to employ makeup waters that are blowdown streams from other plant water systems. Water balance considerations make cooling tower blowdown a particularly desirable spray dryer makeup water source. A series of tests was conducted on the Arapahoe pilot unit to determine the suitability of two simulated cooling tower blowdown (CTB) streams as makeup to the system. The compositions of the two simulated CTB streams are summarized in Table 4.

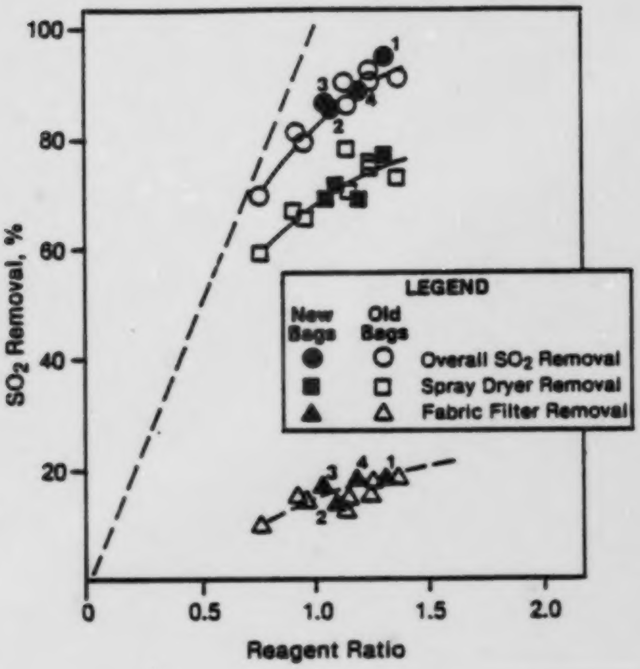

Figure 14. Comparison of 1000 ppm Inlet SO$_2$ Recycle Results with Old and New Bags in Fabric Filter

TABLE 4. NOMINAL COMPOSITIONS OF SIMULATED COOLING TOWER BLOWDOWN WATERS (mg/L)

Component	CTB No. 1	CTB No. 2*
Ca^{++}	1160	410
Na^{+}	480	1830
Mg^{++}	-	53
$SO_4^{=}$	1000	5080
Cl^{-}	2050	170

*Includes 10 ppm AMP, HEDP, or polyacrylate scale inhibitors for some runs.

CTB No. 1 was chosen because it represents the sulfate levels of a gypsum limited, acid treated system. Originally of interest in this water, then, was the sulfate content of 1000 ppm. This sulfate content has generally been accepted as being too high for use as slaking water. Poor slaked lime quality has been reported to result from the use of such waters in slaking.[3] In order to evaluate these impacts for a spray dryer based FGD system, tests were conducted with CTB No. 1 used as slaking water, makeup to the recycle/lime slurry mix tank, and temperature control water. In one test, CTB No. 1 was not used for slaking, but still used in the mix tank and for temperature control. The results of these tests are compared to baseline 1000 ppm inlet SO_2 recycle tests in Figure 15. The results in Figure 15 are on the surface quite surprising. All of the tests using CTB No. 1 show elevated SO_2 removal performance both in the fabric filter and overall.

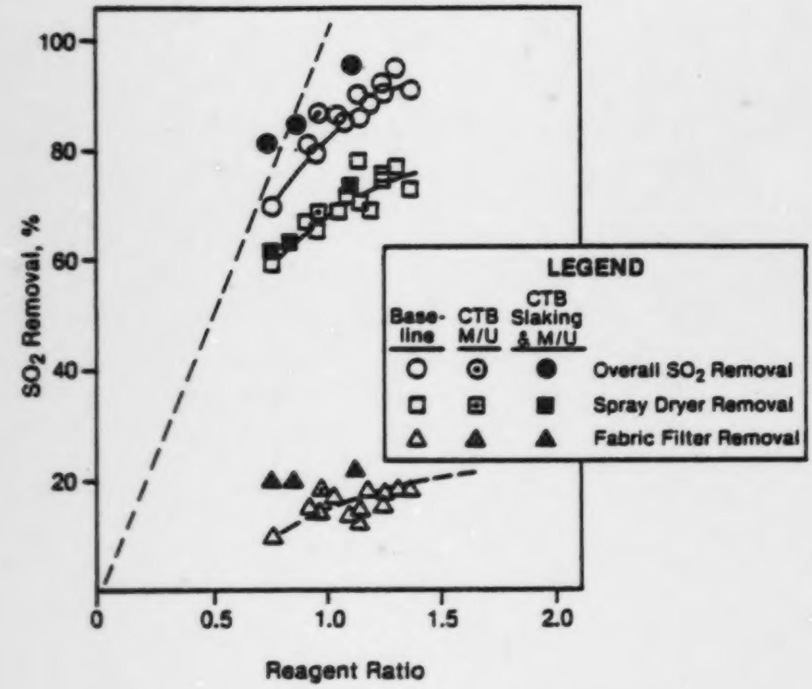

Figure 15. 1000 ppm Inlet SO_2 Recycle Results for Baseline Conditions When Using Simulated Cooling Tower Blowdown No. 1 for Slaking and Makeup Water

A further examination of simulated CTB No. 1 shows that the nominal chloride level is quite high, at 2050 ppm. Although not yet confirmed by chemical analyses, material balance calculations indicate that for the three tests where CTB No. 1 was used as slaking, mixing, and temperature control water, steady-state atomizer feed chloride levels were approximately 5000 ppm, and the fabric filter solids had approximately a 1 percent chloride content. These levels are very near those reported for substantial SO_2 removal enhancement by others.[4,5,6] The enhancement effect may be caused by the deliquescent properties of $CaCl_2$, resulting in a lengthened second drying phase time period in the spray dryer, and significantly increased residual moisture levels in the solids collected on the fabric filter. Both impacts would tend to improve SO_2 removal performance.

The results in Figure 15 appear to substantiate a chloride enhancement effect. The three tests involving CTB No. 1 for slaking, mixing, and temperature control show dramatically increased fabric filter SO_2 removal contribution. The one test employing CTB No. 1 as mixing and temperature control water only shows a somewhat reduced effect. This is expected because the use of service water for slaking tends to dilute the steady-state chloride level. While the presence of this chloride enhancement effect may mask the impact of high sulfate levels on lime slaking, it does appear that any detrimental efect of 1000 ppm sulfate levels is small compared to the potential benefits of elevated chloride levels. One consideration in evaluating these results, though, is that the chloride level in CTB No. 1 is artificially high, significantly higher than might be seen in a typical CTB water. The high chloride content in CTB No. 1 is a result of adding $CaCl_2$ to produce a gypsum relative saturation of near 1.0 for the 1000 ppm sulfate level simulated blowdown water.

The second simulated cooling tower blowdown water in Table 4, CTB No. 2, accurately represents a side-stream softened cooling water system. This water contains a significant amount of sodium as the result of softening, and consequently low calcium and high sulfate levels. Four tests were conducted with this simulated CTB; one with the constituents in Table 4 alone, and one each with 10 ppm as active ingredient of HEDP, AMP, and polyacrylate scale inhibitor, respectively, added. These tests were conducted because most cooling towers are treated with scale inhibitors. Scale inhibitors are intended to delay the precipitation of limited solubility species on heat exchanger surfaces. However, precipitation of these species is the very operation occurring as water evaporates from the droplets in a spray dryer. Although the dosage of scale inhibitors encountered in most cooling tower blowdowns is relatively low (<10 ppm), the net effect of such inhibitors on spray dryer SO_2 removal or solids drying performance was unknown. The results of these tests are presented in Figure 16. The results show that, in general, SO_2 removal across the spray dryer and for the overall system is enhanced by the use of CTB No. 2 as mixing and temperature control water (slaking water use was not tested). The individual tests are not identified because there is some scatter in the four tests conducted, and individual points are not sufficient to rank the magnitude of enhancement resulting from the use of the three generic scale inhibitors. What is important in Figure 16 is that for all of the four tests, overall system performance is enhanced by the use of the simulated CTB as makeup.

Based on the encouraging test results presented in Figures 15 and 16 for the use of simulated cooling tower blowdowns as system makeup water, a number of additional tests using other simulated as well as actual CTB waters are planned for the coming months.

SUMMARY AND CONCLUSIONS

A number of significant conclusions can be made at this point in the pilot test program. These include:

- Spray dryer operability and SO_2 removal performance both are improved by operation in the recycle rather than once-through mode. The improvements appear to be related to increased solids content in the slurry fed to the atomizer as well as to the other benefits of recycling in improving sorbent utilization.

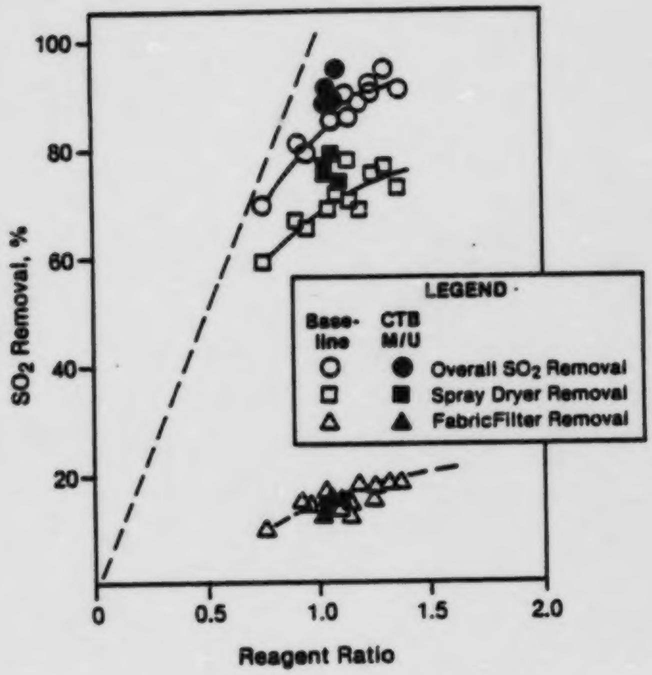

Figure 16. 1000 ppm Inlet SO$_2$ Recycle Results for Baseline Conditions and When Using Simulated Cooling Tower Blowdown No. 2 for Makeup Water

- Operation of a fabric filter downstream of a spray dryer appears to be acceptable in terms of bag fabric impacts, and may actually be beneficial with respect to pressure drop performance. The one area of concern is that of corrosion of compartment walls and mild steel bag caps in cold spots within the compartments. Corrosion problems can most likely be avoided through the use of adequate insulation thicknesses and careful design and installation practices to avoid localized cold spots.

- For low spray dryer inlet SO$_2$ concentrations, SO$_2$ removal performance and lime reagent utilization remains quite acceptable in spite of very low dryer inlet temperatures (as low as 210°F).

- Cooling tower blowdown waste streams available in most power plants appear to be an acceptable and in some cases beneficial source of makeup water to a spray dryer based FGD system. Typical scale inhibitors present in these cooling tower blowdown streams do not appear to adversely affect SO$_2$ removal or solids drying performance.

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APPENDIX

CONVERSION OF BRITISH ENGINEERING (ENGLISH) UNITS TO SI UNITS

To Obtain	From	Multiply by
ng/J	lb/10^6 Btu	430
m	ft	0.3048
m^2	ft^2	0.0929
m^3	ft^3	0.0283
mg/nm^3	gr/dscf	1.83
°C	°F	$t_c = 5/9(t_f-32)$

SESSION 10, PART II: DRY FGD: FULL SCALE INSTALLATIONS

Chairman: Richard G. Rhudy
Electric Power Research Institute
Palo Alto, CA

FIELD EVALUATION OF A UTILITY DRY SCRUBBING SYSTEM

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ABSTRACT

This program, co-funded by the U.S. Environmental Protection Agency and the Electric Power Research Institute, has resulted in an evaluation of a full-scale utility spray dryer/baghouse dry FGD system. The system is installed at the Northern States Power Company's Riverside Station and treats flue gas from a nominal 100 MW of coal-fired power generation. This has been the first independent evaluation of a full-scale spray dryer/baghouse system.

For the test program, two different coals were used as boiler fuels. One coal was a subbituminous coal and coke mixture with a nominal 1.2 percent sulfur content. The second was a 3.4 percent sulfur Illinois bituminous coal.

During the test program, SO_2 removal, particulate emissions, sulfuric acid removal, and extensive process data were recorded. The test program was conducted from July to October 1983, so only preliminary results are presented. Low sulfur coal tests indicated up to 90 percent SO_2 removal was achievable in the short term with slightly sub-stoichiometric amounts of lime addition. A similar removal was achieved in short term tests with high sulfur coal at reagent ratios of 1.3 to 1.4. Calcium chloride addition was found to reduce the lime addition requirements for high sulfur tests by approximately 25 percent.

INTRODUCTION

This paper presents preliminary results from a program entitled "Field Evaluation of a Utility Dry Scrubbing System." The objective of the program has been to acquire performance data on an operating, utility-scale, spray-dryer-based, dry FGD system. The system was evaluated primarily to determine SO_2 and particulate removal performance and lime reagent consumption. The system chosen for evaluation is the Joy/Niro Demonstration Unit located at the Northern States Power Company (NSP) Riverside Station in Minneapolis, Minnesota. The Riverside system was chosen for this program because it is the first lime-based system in operation using a full-size (46 ft* diameter) spray dryer module. Testing was conducted with both low and high sulfur boiler fuels. The program is being conducted for the Environmental Protection Agency, Industrial Environmental Research Laboratory in Research Triangle Park, and for the Electric Power Research Institute under a cooperative funding arrangement.

This test program was needed for several reasons. First, a significant number of spray-dryer-based dry FGD systems have been sold to the utility industry. At least 17 systems representing approximately 6,800 MW of electric generating capacity have been sold to the utility industry. No independent evaluation of the technology is available to aid utilities purchasing dry FGD systems, since virtually all of the performance data have come from pilot- or demonstration-scale units operated by system vendors. Several full-scale systems are coming on-line, but little performance data on these systems have been published. A second consideration is that most of the new systems and all of the future systems must meet the 1978 New Source Performance Standards for coal-fired utility boilers, which call for 70 to 90 percent SO_2 removal on a 30-day rolling average basis and particulate emissions of 0.03 $lb/10^6$ Btu heat input or less. Few data exist to confirm the ability of full size spray dryer-based dry FGD systems to achieve these levels of performance, particularly for high sulfur coal requiring 90 percent SO_2 removal. Finally, the spray dryer-based systems sold to date have been justified using economics based on vendor guarantees for lime consumption. However, no independent evaluation of lime consumption on an operating full-scale system has been conducted. The program described in this paper was designed to collect the type of data that can help determine the suitability of this technology for utility application.

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*British engineering units rather than SI units are used in this paper because of customary usage in the electric power industry. An appendix provides appropriate conversion factors.

The on-site test program was conducted from July 11 to October 8, 1983. Because little time has been available for data evaluation, this paper does not include a complete evaluation of results. Instead, the project is described, system operation is discussed, and preliminary performance results are presented.

PROJECT DESCRIPTION

The project description includes a description of the Riverside Station and FGD system, a summary of the test program, and a discussion of the limitations of the Riverside system and how they have affected this evaluation.

SITE DESCRIPTION

The Riverside Generating Station, operated by Northern States Power Company, is located on the east bank of the Mississippi River in northeast Minneapolis. Some parts of the facility are well over 50 years old, and the two units of interest on this project, No. 6 and No. 7, began operation in 1949. The combined generating power of Units 6 and 7 is rated at 98 MW. However, the pulverized coal, wall-fired units were originally designed to fire an eastern bituminous coal. Recently, the units have fired a western (Sarpy Creek) subbituminous coal. A small amount (10 to 15 percent) of high sulfur coke is added to the subbituminous coal to improve its firing properties. The units can still be fired with high sulfur bituminous coal and, in fact, were so fueled for five weeks of this test program.

Units 6 and 7 are used as peaking units in the NSP system. During the cooler months of the year, the units are rarely operated. During the summer months, the units are operated at up to 90 MW during daytime hours and at a minimum load during the night. The units are normally banked over the weekend during the summer.

In 1980, a full-scale, Joy/Niro, spray dryer/fabric filter FGD system was installed to treat the combined flue gas from the two units. The fabric filter was actually purchased by NSP because of the inability of existing ESP collectors to efficiently collect the ash from the western coal. The spray dryer system was installed by the Joy/Niro joint venture under a cooperative agreement with the utility to serve as a full-scale demonstration of the capabilities of their dry FGD system. Figure 1 is a simplified flow diagram for the system.

The spray dryer is a 46-ft diameter vessel, with flue gas introduced both above the atomized spray in a roof gas disperser and below the atomized spray in a central gas disperser. A rotary atomizer is used, currently employing a 700 hp drive motor. The spray dryer was sized to treat flue gas corresponding to a 70 MW boiler load. This reduced sizing allowed the capability to test the system at greater than design flow rates. It should be noted that, because Units 6 and 7 are over 30 years old, the flue gas flow rate at 70 MW is equivalent to the flue gas rate from approximately a 100 MW new unit. A new unit would experience much less air inleakage and operate at a much lower net plant heat rate than these units. The downstream fabric filter contains 12 compartments, in 2 rows of 6 compartments each. Because the fabric filter was sized to treat hot flue gas, it is actually oversized when the spray dryer is in operation because of the flue gas volume shrinkage which results from the reduced spray dryer outlet temperature.

Pebble lime reagent is slaked in a Joy/Denver attrition slaker. A Joy/Denver ball mill is also available for lime slaking. Milk of lime, dilution water, and recycle solids are added to a mix tank at rates determined by a Honeywell process control computer. The mix tank effluent is pumped to a separate atomizer feed tank. From the atomizer feed tank, slurry is pumped to a head tank at the top of the spray dryer. A pinch-type control valve regulates the flow of slurry to the atomizer to maintain either a constant spray dryer outlet temperature or a constant approach to adiabatic saturation. When the system is operated in an SO_2 removal control mode, the Honeywell process control computer calculates the amount of lime which must be added upstream at the mix tank in order to achieve the desired SO_2 removal. Recycle material is added at the mix tank at a rate required to bring the mix tank solids level up to a set point, normally 35 weight percent solids. The recycle solids are collected from the spray dryer bottom dropout and largely supplemented by a portion of the fabric filter catch.

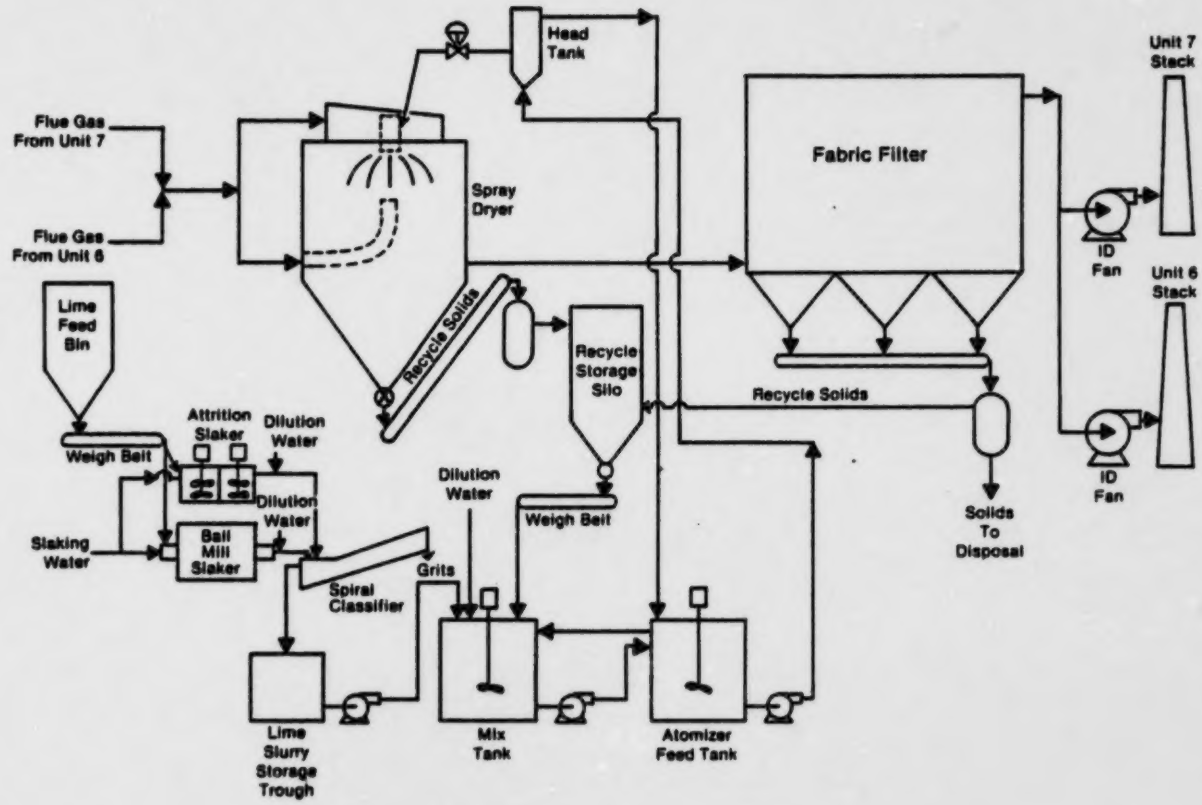

Figure 1. Flow Diagram for Riverside Dry FGD System

The atomizer feed scheme tested is now an obsolete design. The mix tank and atomizer feed tanks used at Riverside each have about 45 minutes of residence time. Thus the system is relatively slow to respond to changes in spray dryer inlet flue gas conditions (i.e., changing SO_2 concentration, inlet temperature, or wet bulb temperature). Current Joy/Niro system designs call for preparing milk of lime slurry and recycle slurry in separate tanks and then mixing them in a tank with a relatively short residence time. This should result in a quicker response to process changes.

TESTING APPROACH

As described in the introduction, a primary objective of the program has been to quantify SO_2 removal by the system. A continuous emission monitoring system (CEMS) has been installed to quantify SO_2 removal. The CEMS includes a DuPont Model 460 two-point extractive SO_2 monitor and a Thermox O_2 monitor sampling the spray dryer inlet and spray dryer outlet ducts, and a Lear Siegler SM810 in-situ point-type SO_2 analyzer installed in a short run of duct at the fabric filter outlet. A second Thermox O_2 analyzer is mounted on the duct exactly opposite the Lear Siegler monitor.

Other than SO_2 removal data, lime consumption and other important process parameters were recorded as hourly averages for each test day. Lime consumption was measured primarily by determining the lime content of the milk of lime slurry introduced to the atomizer feed mix tank. The flow rate of this slurry was continuously measured with a magnetic flow meter and recorded by the Honeywell process control system computer. Other methods of lime consumption measurement have included continuous quicklime weigh belt rate measurements, recording of daily quicklime truckload deliveries, and determination of the lime content and flow rate of the actual atomizer feed slurry. Energy balances have been used to confirm agreement between slurry feed rate and flue gas flow measurements.

Other than quantifying SO_2 removal and lime consumption performance for the spray dryer/baghouse system, determination of particulate removal performance for the system was also a primary objective. This performance was determined by manual sampling of flue gas streams for particulate concentrations, using EPA methods. Particulate loadings were measured at the spray dryer inlet, spray dryer outlet, and fabric filter outlet locations.

TEST PLAN

The test schedule is summarized in Table 1. The schedule shows four different sets of conditions with low sulfur, Sarpy Creek coal/coke blend, and three different sets with a high sulfur Peabody Illinois coal. The Sarpy Creek coal/coke blend has a nominal sulfur content of 1.1 to 1.2 percent and a heating value of 9300 Btu/lb. New Source Performance Standards for utility boilers would require 75 to 80 percent SO_2 removal for boilers firing a fuel with this sulfur and heating value. In some localities, state or local regulations might require as high as 90 percent SO_2 removal with this fuel. Consequently, low sulfur tests were conducted at both 75 percent and 90 percent target SO_2 removal levels. Additionally, some tests were scheduled at a fabric filter air-to-cloth ratio higher than the normal value of 2:1 and with the

TABLE 1. TEST SCHEDULE AND TARGET SYSTEM OPERATING CONDITIONS[a]

Fuel	SO2 Removal Level, %	Fabric Filter A/C Ratio, cfm/ft2	Slaker Type
Sarpy Creek Coal/Coke	75	2:1	Attrition
Sarpy Creek Coal/Coke	75	2.2:1	Attrition
Sarpy Creek Coal/Coke	90	2:1	Attrition
Sarpy Creek Coal/Coke	90	2:1	Ball Mill
Illinois Coal	90	2:1	Ball Mill
Illinois Coal	90	2:1	Attrition
Illinois Coal[b]	90	2:1	Attrition

[a] All tests planned to be conducted at an 18°F approach to adiabatic saturation, 35 weight percent solids in the atomizer feed slurry, a 70 MW daytime boiler load, and with a once-per-hour baghouse cleaning frequency.

[b] Calcium Chloride Addition Tests

ball mill slaker rather than attrition slaker being used to prepare the lime slurries.

The test at the higher air-to-cloth ratio was originally intended to be conducted at an air-to-cloth ratio near 2.7:1. However, the air-to-cloth ratio is increased at a constant boiler load by taking individual compartments off-line. After taking compartments off-line to raise the gross air-to-cloth ratio to near 2.7:1, only 6 of the 12 compartments remained on-line at a 70 MW boiler load. The combined effects of taking an operating compartment off-line for cleaning and reverse gas flow during cleaning momentarily sends the effective air-to-cloth ratio to 3.5:1 to 4:1. In a unit with more compartments on-line, the effects of cleaning one compartment in increasing the effective air-to-cloth ratio are much smaller. Additionally, in the event of a temporary loss of slurry flow to the atomizer while a compartment is cleaning, the increased flue gas volume due to the much higher spray dryer outlet temperature would result in a further increase in the effective air-to-cloth ratio to well over 4:1. When operated at such a high effective air-to-cloth ratio, the pressure drop across the fabric filter is increased to several times its normal value. Such a combination of events did occur on the first day of the high air-to-cloth ratio tests, and the resulting momentary high pressure drop caused the control system to bypass the fabric filter. In order to provide a margin of safety, one more compartment was put on-line. This lowered the gross air-to-cloth ratio to 2.2:1 to 2.3:1, which is only 10 to 15 percent higher than baseline conditions.

The fourth set of conditions for the low sulfur tests (see Table 1) was to have included ball mill rather than attrition slaked lime. However, ball mill operation during this test program was generally not successful. The reasons for the ball mill operating problems are further discussed under Operational Results. Because of these problems, less than one day of system operation was observed using ball mill slaked lime.

Two sets of conditions were tested with the high sulfur coal. The coal was an Illinois No. 6 coal with a nominal 3.4 percent sulfur content and 10,800 Btu/lb heating value. Current New Source Performance Standards for utility boilers require 90 percent SO_2 removal when coal of this sulfur content and heating value is burned. Consequently, only a 90 percent target SO_2 removal was tested with this high sulfur coal. The tests included baseline conditions of 90 percent removal with attrition-slaked lime and a second test run employing calcium chloride addition for lime utilization enhancement. Chloride addition has been reported previously to enhance lime utilization in spray dryer/baghouse FGD systems[1,2], but this is the first test of chloride addition at a full-scale utility installation. The enhancement effect is thought to occur because of the deliquescent properties of calcium chloride. This effect is further discussed in the System Performance portion of this paper.

Originally, ball mill slaker tests and high air-to-cloth ratio tests were planned for the high sulfur coal portion of the test program. However, due to the ball mill slaker operating problems previously mentioned and the previously discussed constraints on gross air-to-cloth ratio, only the last two tests listed in Table 1 were conducted during the high sulfur test period.

SYSTEM LIMITATIONS

Several system limitations combined to restrict the amount and the type of data that could be collected. First, as mentioned earlier, Riverside Units 6 and 7 are peaking units. As such, they are rarely operated in the winter and only operate part-time during July to October. This part-time operation involves unit loads of 70 to 90 MW during weekday daylight hours, minimum load (30 to 50 MW) overnight during the week, and banking the boilers over the weekend. Figure 2 is an illustration of the combined Units 6 and 7 loads during a typical week. Therefore, although the FGD system was operated at desired SO_2 removal levels 24 hours per day, only about 12 hours per weekday of near full-load operation were available for evaluating FGD system performance. At the beginning of each 12-hour full-load period, the FGD system generally goes through a transient period due to a large increase in boiler load. On Mondays the unit must undergo a cold start-up. Although this cycling provides a severe test of the capabilities of the system, it reduces the period of steady state operation at the desired SO_2 removal level over which lime reagent consumption can be measured.

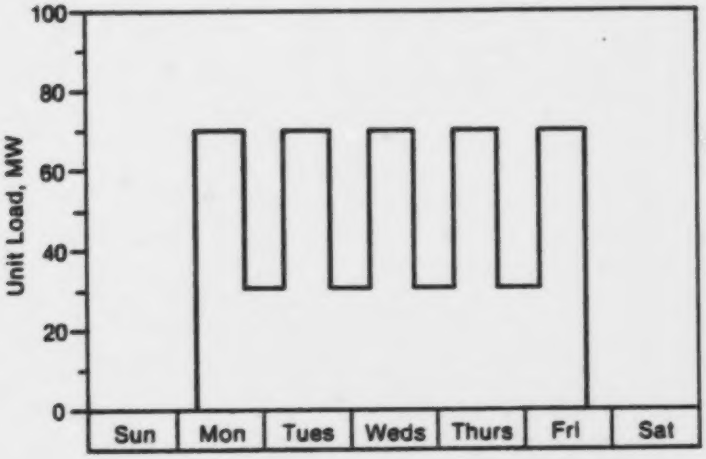

Figure 2. Illustration of Typical Unit Load During Test Program

Additionally, the Riverside system was the first utility-scale system designed and built by the Joy/Niro joint venture and was built as a demonstration unit. As the first unit built, the Riverside system has provided the opportunity to refine and modify design features for subsequent systems.

Thus, some specifics of the Riverside system are different from what will be found in later designs. An example of this is the slurry feed system design, which has been modified for subsequent system designs to provide quicker response to transients such as load changes. At Riverside, lime slurry, makeup water, and recycle solids are mixed in a single tank. The lime slurry addition rate is based on SO_2 removal (for the SO_2 removal control mode of operation), the makeup water rate is controlled to maintain level in the mix tank, and the recycle solids rate is controlled to achieve a set weight percent solids (normally 35 percent solids). The mix tank has a residence time of approximately 45 minutes. This slurry mixture then overflows to an atomizer feed tank with a similar residence time. When the inputs to the mix tank change composition abruptly (such as when the unit load increases), approximately 3 or 4 hours are required before the feed system stabilizes near steady state for the new conditions. Recently designed Joy/Niro systems have a faster response slurry feed system. In these systems, recycle solids and makeup water are mixed in a recycle slurry tank to a consistency of approximately 50 to 60 weight percent solids. The lime and recycle slurries are then mixed in a feed tank with a relatively short residence time of one-half hour or less. Because only this feed tank residence time impacts the system response time, this feed system might stabilize after an abrupt change of operating conditions in one hour or less.

The impacts of the slurry feed preparation system design are particularly important at the Riverside Station, as the normal station operation causes significant load changes at least twice per day. In fact, on some days during this test program the NSP dispatcher called for the unit load to be varied between 70 MW and 90 MW throughout the day. On these days, the unit never operated at one load long enough for the feed system to stabilize. On most other days, even if the load was steady all day, only 8 or 9 hours of the 12 hours of full load operation actually represented steady state conditions.

Other aspects of the Riverside system's status as a demonstration unit affected the results of this program. For example, the system contains only one spray dryer module, while most utility systems will be multiple module systems. In a multiple-module system, equipment problems which affect one individual module have a smaller impact on overall system performance. Being a one-module system, equipment problems tended to cause the entire system to have to be shut down, or operated at conditions other than those desired. Even considering that the smaller Riverside system only warrants a single dryer vessel, some individual components of the system have not been installed with the redundancy that the vendor would likely install in a commercial system. These considerations have had a detrimental effect on both the amount of system downtime and the number of process-equipment-related upsets during the test program.

A final consideration which has affected the test program involves the recent history of the Riverside station. For nearly the first two years of operation, the FGD system was used as a full-scale demonstration and testing unit by the process vendors. During this time the Joy/Niro joint venture had responsibility for the operation of the system, even though NSP actually provided operating personnel. Within the year prior to the test program, NSP had assumed responsibility for the operation of the FGD system. Immediately prior to the test program, the Units 6 and 7 boilers and FGD system were off-line for much of the winter and spring, as NSP does not need power production from these units during this time. During this long period of downtime, normal personnel turnover (promotions, retirement, transfers, etc.) resulted in a number of new operators rotating into the FGD system operating staff. Early in the test program, then, the FGD system was being operated for the first time in several months with a staff of operators having little previous experience with the FGD system. Consequently, a lot of test days early in the program were not productive as far as representing steady state operation at test conditions. As the test program continued, successful test days represented a higher fraction of potential days of operation.

Another consideration here is that, after the Joy/Niro joint venture turned over the dry FGD system to the utility, and prior to conducting this test program, NSP only operated the system at 50 to 60 percent SO_2 removal levels. For such moderate SO_2 removal levels, they had previously operated the system under very conservative conditions (i.e., higher spray dryer outlet temperatures). Thus, operation at 75 or 90 percent SO_2 removal, with a 20°F approach to adiabatic saturation, was not a normal operating mode immediately prior to this test program. Early in the program, the operators tended to revert to conservative operation (i.e., higher spray dryer outlet temperature) during any minor upset, such as soot blowing in the boilers. This would move the operation away from the desired conditions and would preclude acquiring desired steady state operating data. As the program continued though, this occurred much less frequently as the operators became more comfortable with operating at test conditions.

Considering the previous discussions, it was not realistic to report availability of the system, as the availability of the Riverside system would tell little about that of a commercial utility, multi-module, dry FGD system on a new base-loaded boiler. The combined effects of weekly cold startups, frequent load changes, little redundancy, and a somewhat undertrained operating staff at Riverside do little to promote a fair assessment of the potential availability of a commercial system.

However, the general operation of the system was closely observed during the test program. Much of the downtime or off-condition operating time was due to problems specific to the Riverside system. Others appear to be more generic to dry FGD systems. These more generic problems are discussed in this paper, as they are more likely to occur in other systems.

RESULTS

The results of the program are divided into two areas: Operational Results, which includes a qualitative discussion of the operation of the system during the test program; and System Performance, which includes preliminary SO_2 removal, lime consumption, and particulate and sulfuric acid removal data.

OPERATIONAL RESULTS

In general, the equipment that comprises the basis of the dry FGD system, the spray dryer and baghouse, were relatively trouble-free throughout the program. At the conditions tested, the spray dryer did not show evidence of potential problems such as wheel nozzle pluggage, excessive buildup of solids on the walls, or formation of wet solids within the dryer. Some atomizer problems were observed, but most of these appeared to be caused by circumstances specific to the situation at Riverside rather than being generic to the Joy/Niro system. These problems will be discussed further later in this section.

The baghouse also operated well, with no significant bag/fabric related problems being observed. In this short-term test though, long-term effects such as bag life or compartment wall corrosion rates could not be evaluated.

Some problems were observed in four specific areas--the slurry feed system, the ash handling system, the ball mill slaker, and in atomizer protection. The system vendors may have addressed these problems in system designs subsequent to Riverside, but the problems could be encountered in virtually any spray-dryer-based, dry FGD system. Each of these areas is discussed below.

Slurry Feed System--In a recycle lime system, lime slurries containing up to 25 percent solids and recycle/lime slurries of 30 to 40 percent solids are commonly encountered. When dealing with slurries with a high solids content and high viscosity, problems such as plugging of pump suction lines, solids buildup on tank walls, plugging of in-line screens used to remove oversize material, and loss of flow when switching pumps are commonly encountered. Such problems were encountered often at the Riverside system. Years of operation of wet lime/limestone FGD systems have established means of dealing with such problems: always keep slurry in tanks well agitated, never allow slurry levels to drop to the agitator blade level, never allow slurry to stand in filled lines, allow for on-line changing of plugged or partially plugged screens, etc. These same solutions will need to be applied in spray drying systems.

The quantities of these slurries that must be dealt with in a spray dryer system are much smaller than what would be encountered in a wet system. At Riverside, typical atomizer feed slurry rates are 150 to 200 gpm. In a similarly sized limestone wet FGD system, the slurry recirculation rate could be as high as 40,000 gpm. The point to be made is that while some of the slurry handling problems of a wet FGD system may still be encountered in a spray dryer system, they will occur on a greatly reduced size of equipment. This should make both problem solving and routine maintenance easier. Another important point to be noted is that no chemical scaling tendencies were observed at Riverside.

Ash Handling System--In comparing the wet versus dry FGD systems the spray dryer system has a slurry feed system that deals with much lower flow rates for a given unit capacity, but the quantities of dry ash and FGD by-products that must be moved around the system are substantially greater than for a comparably sized particulate collection device/wet scrubber system.

For example, at Riverside the amount of solids entrained in the flue gas at the fabric filter inlet (spray dryer outlet) is 3 to 5 times that in the spray dryer inlet gas. For low sulfur coal operation only a small portion of this increase is attributable to the lime reagent introduced and SO_2 removed in the spray dryer. The majority is material being recycled from the fabric filter catch to the spray dryer. Thus, the rate at which solids are removed from the fabric filter hoppers and transported to recycle storage bins can be several times greater than the net rate of waste product actually leaving the system.

Specifically, for low sulfur coal, 75 percent SO_2 removal, and the unit operating at 80 MW, the steady state rate of fly ash and spray dryer by-products leaving the system amounted to 7 to 8 tons/hr. However, due to the effects of recycle, approximately 20 tons/hr were actually being collected by the baghouse. The excess between that going to the disposal silo and the 20 tons/hr being collected, or 12 to 13 tons/hr, was being recycled to the slurry feed system. In this example, roughly three times the quantity of ash and spray dryer solids leaving the system must be continually removed from the baghouse hoppers, and roughly twice the quantity leaving the system must be continually transported to the recycle storage bin.

At Riverside, solids collected in the baghouse hoppers and at the bottom of the spray dryer are intermittently dumped onto mechanical conveyors through motor-driven tipping valves. One conveyor each handles ash from the six compartments on the north and south sides of the baghouse, respectively, while a third conveyor handles spray dryer bottom solids. These conveyors empty into surge bins mounted above air conveying blow pots. The blow pots cycle through the following five consecutive processes: filling from the surge bin, isolation from the surge bin, pressurization, pneumatic conveying to either the recycle or disposal storage silos, and then depressurization. The recycle silo empties through a rotary valve onto a weigh belt which meters recycle material into the mix tank.

The problems with solids handling which most frequently occurred at Riverside involved the baghouse mechanical conveyors, blow pots, and the recycle bin rotary valve. The problems with the mechanical conveyors appeared to be related to marginal capacity. The north or south conveyors frequently tripped out because their drive motors were overloaded. Momentary overloading of the conveyors is aggravated by the nature of the operation of the baghouse, where compartments clean intermittently throughout each hour. That is, the average rate at which the baghouse collects material might be 20 tons/hr, as in the example above. Each conveyor would see an average of 10 tons/hr if the same number of compartments were on-line on both sides. However, this average includes short periods of high solids flow, immediately after a cleaning compartment begins reverse gas flow, and periods of low solids flow rate between compartment cleaning periods. Thus, one conveyor may see instantaneous rates of several times 10 tons/hr. This problem was apparently corrected after this test program was completed by a modification which meters the rate at which ash/spray dryer by-products are emptied from the hoppers onto the conveyor.

In addition, the conveyors require occasional maintenance, such as replacement of failed drive motors and bearings. If one conveyor must be off-line for a very long time, then all six of the compartments emptied by that conveyor must be taken off-line once the hoppers below those compartments are full of ash. At Riverside, failure of one conveyor can cause the boiler load to be limited to that which can be handled by the six compartments remaining on-line. Because the rate at which solids are collected in a fabric filter downstream of a spray dryer is greatly increased, unless the hoppers are increased in size compared to conventional particulate control configurations, such a conveyor failure can very quickly limit unit load.

Failure of a blow pot can have a similar effect, as only a small surge bin separates each conveyor from a blow pot. NSP has attempted to provide some redundancy in blow pots, but in an instance where only one blow pot is available to accept solids from a conveyor, a blow pot outage can cause the conveyor to be shut down as well. Common blow pot failures observed included plugging, leaking isolation valves, and loss of air pressure to the pneumatically operated valves.

A third piece of solids handling equipment which rapidly affected the operation of the system on failure was the rotary valve at the bottom of the recycle solids bin. When this valve failed, often due to a piece of tramp material jamming the blades, the solids content in the atomizer feed slurry immediately began to drop, and the system began to approach once-through lime operation. The large slurry mix tank and atomizer feed tank could be considered a 1.5 hour reservoir of feed slurry, but in order to avoid splashing slurry on tank walls and a resulting buildup of dry solids, the tanks must be continually kept near full. Consequently, these tanks begin to dilute immediately when recycle solids flow is interrupted because makeup water must be added to maintain level.

The solids handling equipment appears to be a key to successful operation of the spray dryer/baghouse system. The individual pieces of equipment must be sized to handle solids rates much higher than is typical for conventional particulate removal alone.

Ball Mill Slaker--As mentioned previously, during this program the ball mill slaker did not operate successfully for any extended period. The feed end of the slaker tended to plug with wet lime solids. While there are several possible reasons why the plugging continually occurred, the actual cause was not identified. The ball mill operation is discussed below.

The ball mill slaker is fed by an independently driven screw feeder inserted directly through the trunion bearing of the mill. Product lime flows out the opposite end, through a cylindrical screen, and into a product collection trough. Figure 3 is a simplified cross section of the ball mill depicting the feed and product ends of the mill. Two things are clear from this simplified illustration: first, vapors produced by the heat of the slaking reaction can only leave the slaker with the product or flow back through the feed end. Secondly, little hydraulic head is available to promote the flow of the viscous slaked lime slurry out the product end of the slaker without immersing the feed screw in slurry.

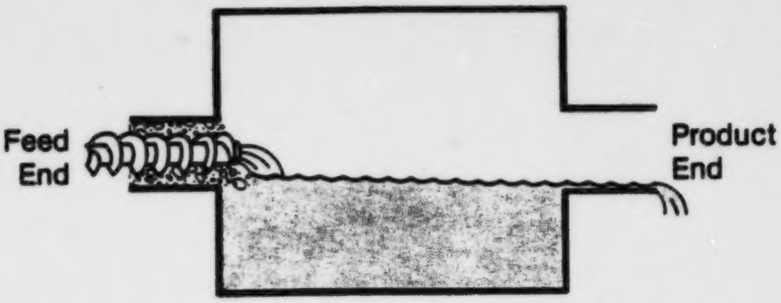

Figure 3. Simplified Cross Section of Ball Mill Slaker

Either of these considerations may have contributed to feed screw pluggage. The ball mill slaker was operated at a very low water-to-lime ratio, resulting in 30 to 35 percent solids in the mill product slurry and a product temperature of 97°C (207°F) or greater. At these conditions, the product slurry is very viscous, having a paste-like consistency. Also, with the product slurry very near or perhaps actually boiling, vapor production rates are high. Both of these effects of low water-to-lime ratio can promote feed screw pluggage. Another consideration is that the ball mill slaker had not been operated for some time prior to this test program, as NSP prefers to operate the attrition slaker at Riverside. The mill had not been completely cleaned out after its last use, and solids buildup around the discharge end undoubtedly contributed to the slurry level in the mill backing up into the screw feeder.

A final problem specific to the Riverside system may also have contributed to feed screw pluggage. The SO_2 removal software in the process control computer tended to adjust the lime slurry makeup rate to the mix tank after a major change in the process (e.g., unit load change) in a classic dampened controller response. That is, after an abrupt process change the lime slurry makeup rate to the mix tank would tend to overshoot, then undershoot, for several cycles, before stabilizing at the required value. During the undershoot portion of the response, the lime slurry feed rates might be completely stopped. As the Riverside system has little lime slurry storage capacity, the slaking rate tended to directly follow the slurry makeup rate. Thus, periods of high slaking rates would be followed abruptly by periods where the screw feeder would shut down. Shutting down the screw feeder momentarily several times in succession undoubtably contributed to the plugging problems.

The ball mill slaker problems at Riverside appear to be somewhat site-specific as ball mill slakers have operated successfully elsewhere. In retrospect, the slaker might have run more successfully at a higher water-to-lime ratio, resulting in a less viscous product slurry and less vapor release. However, in the attempts to run the ball mill slaker during the high sulfur tests in particular, the slaking water piping size did not allow operation at higher water-to-lime ratio at the lime slaking rates required. During the high sulfur tests, lime slaking rates averaged approximately four times that required for the normal low sulfur fuel at similar unit load and percent SO_2 removal conditions.

Atomizer Problems--Although the atomizer motor, gearbox, and nozzle wheel were generally trouble-free, on several occasions a problem which could result in atomizer damage was observed.

Two scenarios for potential damage were observed. One occurred when the unit was forced to run from the basic, or less sophisticated, control station of the computer control system while the more sophisticated supervisory station was undergoing repair. The control software at Riverside does not have full interlock protection for the atomizer when running from the basic station. Interlocks are software which automatically shut down the atomizer when given inputs that are indicators of problems which might result in damage to the atomizer. While running in this mode, a minor problem involving loose wires to the atomizer oil circulating pump occurred, intermittently shutting the pump off. However, the basic station only gave the control operator an alarm rather than shutting the atomizer down automatically. The atomizer continued to run for several minutes without oil circulation and sustained gearbox damage.

This incident may be very site-specific to the Riverside system due to the limitations discussed earlier. First, Joy/Niro has reported that their commercial systems have full atomizer interlock protection in both the supervisory and basic computer control stations. Second, an operating staff with more experience on a spray dryer system might have shut down the atomizer more quickly when given an indication of a loss of oil flow and would not have attempted to restart the machine prior to remedial action.

The second scenario which could result in atomizer damage was observed on more than one occasion. This incident involved feeding slurry to the atomizer wheel when it was not rotating. Since the non-rotating wheel has a much lower hydraulic capacity than a rotating wheel, slurry fed to the standing wheel tends to overflow the wheel and can flow up the spindle to which the wheel is attached and enter the atomizer oil system. In such instances the oil sump can be immediately emptied and flushed to avoid damage, but if the atomizer is operated before cleaning, the slurry in the oil can eventually cause gearbox damage.

Such an instance of feeding slurry to a standing wheel can occur from operator error in establishing flow to the atomizer when it is not in operation and/or starting the atomizer from the basic station without full interlock protection. As for the previous discussion, such an occurrence is much less likely for a commercial Joy/Niro system, as even the basic station would have full interlock protection. The interlock system would normally close a slurry feed block valve before the atomizer shut down and would not allow this valve to open until the atomizer had returned to operation at full speed.

However, even with a properly operating interlock system, the previously described problem could occur. One such event occurred during this test program. In this instance, during an atomizer trip, valve seat damage prevented the atomizer feed slurry block valve from closing. In this case, all of the interlocks operated properly, but a mechanical problem allowed slurry flow to the standing wheel. An alert operator could catch such a problem very quickly, but it is still likely that slurry would get into the atomizer lube oil unless the flow was stopped immediately.

It is interesting to note that in the example above, the spray dryer was hardly upset by full slurry flow to a standing wheel. Several thousand gallons of slurry were fed in one instance, and the only repercussions were a mess under the dryer vessel where the slurry flowed out of the dryer onto the ground, and a minor buildup of dried solids on the vanes of the central gas disperser just below the wheel.

What is clear from these observations of potential atomizer damage are two conclusions. First, the rotary atomizer on a spray dryer FGD system should always be operated with the full protection of interlock controls. Second, the control operators should be well trained to respect the potential damage which can result to the atomizer during upset conditions.

SYSTEM PERFORMANCE

The results of SO_2 removal and lime consumption measurements during this test program are summarized in Tables 2 and 3. Table 2 summarizes the low sulfur SO_2 removal results, while Table 3 summarizes those for the high sulfur test period.

For several reasons, the SO_2 removal results from the program cannot be expressed as 30-day rolling averages. First, the unit was never operated at one set of conditions for that long a period during the test program. Also, due to the operating characteristics of the peaking boilers, only 8 to 9 hours per day typically represent full load, steady state operation. Thus, the SO_2 removal and lime consumption results represent values measured during steady state unit operation on only a portion of a number of successive days. The fact that these results are the average of a number of short term tests cannot be used to determine whether or not the dry FGD system could sustain these removal levels over a long period of time; instead, they only reflect the fact that the boiler rarely operated at or near full load for a long period of time.

Lime consumption in Tables 2 and 3 is expressed as a reagent ratio. This term is defined as:

$$\text{Reagent Ratio} = \frac{\text{Moles Calcium in Fresh Lime Fed to System}}{\text{Mole } SO_2 \text{ in Inlet Flue Gas}} \tag{1}$$

(This definition corresponds to that of the term "stoichiometric ratio" in many other dry FGD papers.)

It should be noted that the values in Tables 2 and 3 should be considered preliminary values which are based on a number of manual calculations for each set of conditions. Where possible, each value is supported by alternate calculations. For example, lime slurry feed rates are compared to lime weigh belt readings: essentially a calcium balance on the lime slaker. Also, weigh belt readings have been compared against lime truckload delivery inventories. Flue gas flow rates are checked against slurry feed rates by energy balance calculations. In general, the values in Tables 2 and 3 represent the most accurate values available at this time. As more data reduction is completed, these values will be refined but are not expected to be changed substantially.

Table 2 shows that the desired SO_2 removal levels of 75 percent and 90 percent were achieved for the lower sulfur fuel with lower than or right at stoichiometric amounts of lime. This may be attributable to two factors. First, at 35 percent feed slurry solids in low sulfur operation very high recycle rates are possible. This is seen in the recycle ratio values in Table 2. Recycle ratio is defined in this paper as:

$$\text{Recycle Ratio} = \frac{\text{Lb Recycle Material in Atomizer Feed Slurry}}{\text{Lb Fresh Hydrated Lime } [Ca(OH)_2]} \qquad (2)$$

TABLE 2. PRELIMINARY SO_2 REMOVAL RESULTS, LOW SULFUR

Nominal SO_2 Removal, %	Spray Dryer Removal, %	Fabric Filter Contribution, %	Reagent Ratio	Recycle Ratio
75	67 - 69	7 - 9	0.6 - 0.7	11:1 - 14:1
90	80 - 81	9 - 10	0.7 - 0.8	9:1 - 13:1

TABLE 3. PRELIMINARY SO_2 REMOVAL RESULTS, HIGH SULFUR

Nominal SO_2 Removal, %	Spray Dryer Removal, %	Fabric Filter Contribution, %	Reagent Ratio	Recycle Ratio
90	75 - 77	13 - 15	1.3 - 1.4	2:1 - 3:1
90[a]	67 - 69	22 - 24	0.9 - 1.1	3:1 - 4:1

[a] High Chloride Concentration Tests

Since a large fraction of the baghouse catch and all of the spray dryer bottom solids are recycled, from pilot plant results it would be anticipated that high sorbent utilization would be promoted. Additionally, analyses of the Sarpy Creek coal ash are not yet available, but the ash is known to have a high alkaline earth content (CaO, MgO, Na_2O, and K_2O). These alkaline components are believed to have contributed to SO_2 removal.

The results in Table 2 also show the range of SO_2 removal in the spray dryer (S.D.) and fabric filter (F.F.). SO_2 removal in the fabric filter is calculated relative to the spray dryer inlet SO_2 concentration as follows:

$$SO_2\ \text{Removal}_{F.F.} = \frac{[SO_2\ \text{into F.F.}\ -\ SO_2\ \text{out of F.F.}]}{[SO_2\ \text{into S.D.}]} \qquad (3)$$

With this definition, spray dryer removal and fabric filter removal can be summed directly to yield overall removal values. The results show that at 75 percent overall SO_2 removal, fabric filter removal contributes little to the overall removal. This occurs because sorbent utilization is very high in the spray dryer itself. For 90 percent removal, the fabric filter contribution increases somewhat, but in an amount roughly proportional to the increase in overall removal level.

Table 3 summarizes the high sulfur coal test results. For the first set of data, corresponding to normal low chloride operation, an excess of fresh lime is required. Several factors probably contribute to this effect. One may be that, because of the increased lime addition rates, the recycle ratio is greatly reduced relative to low sulfur values in order to maintain the total solids in the slurry at the desired weight percent value. Also, the Illinois coal ash being nonalkaline contributes no alkalinity to the SO_2 removal reactions. For the first set of high sulfur data, the fabric filter contribution to overall SO_2 removal appears to be more important than for low sulfur operation.

Chloride addition has been reported by others to promote increased sorbent utilization in spray-dryer-based FGD systems[1,2]. The benefits are thought to result from the deliquescent properties of calcium chloride, which delay complete drying of the droplets in the spray dryer and result in higher residual moisture levels in the fabric filter solids. The second set of SO_2 removal data in Table 3 corresponds to the addition of calcium chloride at levels which result in a chloride content of 1 percent in the fabric filter solids collected. This chloride level in the fabric filter solids was recommended by the system vendor, Niro Atomizer, as being an optimum value for lime utilization enhancement, based on pilot-scale studies conducted at their Copenhagen, Denmark, test facility. At Riverside, this solids chloride level required a liquid phase chloride concentration of approximately 7,500 ppm in the atomizer feed slurry. For the recycle rates at Riverside, fresh makeup of calcium chloride accounted for about half of this liquid phase content, and the remainder dissolved from the recycle material. The results in Table 3 show that chloride addition significantly reduced the lime reagent ratio requirements to achieve 90 percent removal; the lime requirement was reduced by about 25 percent.

The SO_2 removal results for these high chloride tests indicate increased SO_2 removal across the fabric filter for 90 percent removal overall. This is an indicator that the benefits of high chloride level on residual moisture level in the fabric filter have a greater impact on SO_2 removal than impacts within the spray dryer.

At Riverside, with chloride levels in the fabric filter solids at 1 percent, residual moisture levels increased from just below 1 percent to nearly 2 percent moisture. These moisture levels are still low enough to avoid problems which result from handling wet solids. Also, no buildup of wet solids on the spray dryer walls occurred during testing, and solids collected at the bottom of the spray dryer contained moisture levels below 4 percent. These tests were conducted at an 18°F approach to adiabatic saturation at the dryer outlet, just as were all previous tests.

Material balance calculations indicate that, for a coal such as the Illinois coal fired in the high sulfur tests at Riverside, a chloride content of around 0.3 percent would provide the chloride levels of this test. This would be an uncharacteristically high chloride level for a typical 3.5 percent sulfur coal. However, using published bulk prices, it appears that calcium chloride could be delivered to a typical plant site for approximately $225/ton on a 100 percent $CaCl_2$ basis[3]. For $70/ton lime, delivered, a net savings would result from chloride addition at Riverside if only a 15 percent reduction in lime consumption resulted. The actual saving observed was well above 15 percent. It appears to be economic in this case, disregarding capital cost considerations, to operate at high chloride levels even if virtually all of the chloride must be added as calcium chloride. For a 3.5 percent sulfur coal with a higher chloride content (0.1 percent or better), the economics would likely be improved. Additionally, a makeup water source with a significant chloride content, such as some cooling tower blowdowns, would further improve these economics.

MASS LOADING MEASUREMENT RESULTS

Table 4 presents mass loading results for both the high sulfur and low sulfur test periods. The results show that the spray dryer increases the grain loading at the fabric filter inlet to 3 to 4 times that of the spray dryer inlet value. The data also show that particulate removal levels remained high throughout the test program. Removal efficiencies across the fabric filter varied from 99.95 to over 99.99 percent. The emission levels in Table 4 are expressed as grains per dry standard cubic foot. Calculation of emission levels in pounds per million Btu cannot be completed until the results of coal ultimate analyses are available, but the pounds per million Btu values should be approximately 1.5 times that in grains per dry standard cubic foot. Therefore, emission rates varied between approximately 0.002 $lb/10^6$ Btu and 0.012 $lb/10^6$ Btu, with most values below 0.006 $lb/10^6$ Btu. The few higher values were related to minor bag problems in one compartment. In all cases, particulate emission rates measured were well below the current NSPS level for utility boilers of 0.03 $lb/10^6$ Btu.

TABLE 4. FLUE GAS MASS LOADING SUMMARY

Sampling Location	Mass Loading, gr/dscf	
	Low Sulfur Tests	High Sulfur Tests
Spray Dryer Inlet	3.2 to 4.1	2.8 to 4.1
Spray Dryer Outlet	11.0 to 13.8	14.9 to 17.2
Fabric Filter Outlet	0.001 to 0.003	0.001 to 0.008

SULFURIC ACID MEASUREMENT RESULTS

Flue gas sulfuric acid concentration measurements were also conducted. For the low sulfur tests, no measurable sulfuric acid levels were detected at either the spray dryer inlet or fabric filter outlet. The inability to measure sulfuric acid at the dryer inlet is apparently related to the alkaline nature of the Sarpy Creek coal ash. It is not clear whether the alkaline ash removes all sulfuric acid upstream of the spray dryer, or whether any sulfuric acid present is removed on the ash collected on the heated filter in the sampling train. At the outlet of the fabric filter though, it is clear that no sulfuric acid is present. Measurements could not be made at the spray dryer outlet due to the high grain loading at that point, as the upstream filters tended to plug before an appreciable amount of flue gas could be sampled.

Measureable levels of sulfuric acid were found during the high sulfur test periods. Spray dryer inlet values were measured at 2 to 6 ppm SO_3. Fabric filter outlet values varied from 0.1 to 0.5 ppm. On a limited number of days where the spray dryer inlet and fabric filter outlet SO_3 concentrations were measured simultaneously, removal efficiencies of 90 percent or better across the system were indicated.

SUMMARY AND CONCLUSIONS

Based on preliminary data reduction for the 3-month test program on the NSP Riverside dry FGD system, the following conclusions are apparent:

- In general, the Riverside system ran quite well. None of the problems anticipated for spray dryer systems, such as rotary atomizer wheel pluggage, buildup of wet solids on dryer vessel walls, or wetting of fabric filter bag surfaces during upset conditions, were observed.

- Some problem areas at Riverside appear to be potential sources of problems on other similar dry FGD systems. These include typical problems with mixing and pumping slurries with a high solids content, solids handling equipment which requires continual maintenance, and sometimes inadequate atomizer protection during upset conditions.

- At sulfur levels up to a nominal 3.5 percent, high SO_2 removal efficiencies (90 percent) were readily achievable in the relatively short-term periods of this program. For the low sulfur Sarpy Creek coal/coke mixture, substoichiometric amounts of lime were required even at 90 percent SO_2 removal. This was attributed to the alkaline nature of the Sarpy Creek coal ash. For the high sulfur Illinois coal, 90 percent SO_2 removal required reagent ratios of approximately 1.3 to 1.4 moles lime per mole of inlet SO_2.

- Calcium chloride addition to the atomizer feed slurry to achieve chloride levels of approximately 1 percent in the fabric filter solids catch appeared to be successful in promoting lime utilization. For the high sulfur tests, the lime reagent ratio to achieve 90 percent SO_2 removal was reduced from 1.3 to 1.4 down to a range of 0.9 to 1.1 moles lime per mole of inlet SO_2. This chloride level would correspond to 0.3 percent chloride in a nominal 3.5 percent sulfur coal. Even for a low chloride, high sulfur coal, high chloride levels achieved through calcium chloride addition appear to be cost effective for reducing lime consumption.

- Particulate control efficiencies were high throughout the test program, maintaining outlet grain loadings well below required levels. In spite of baghouse operation within 18°F of the adiabatic saturation temperature and very high baghouse inlet grain loadings, no bag-fabric-related problems were observed and flange-to-flange pressure drop remained acceptably low.

- Based on a limited number of simultaneous measurements during the high sulfur test periods, sulfuric acid removal levels of 90 percent or greater were observed.

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2. Hansen, Svend Keis, et al. "Status of the Joy/Niro Dry FGD System and Its Future Application for the Removal of High Sulfur, High Chloride and NO_x from Flue Gas." Publication No. 83-JPGC-APC-8, The American Society of Mechanical Engineers, New York, NY. October, 1983.

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APPENDIX

CONVERSION OF BRITISH ENGINEERING (ENGLISH) UNITS TO SI UNITS

To Obtain	From	Multiply by (or use equation)
ng/J	lb/10^6 Btu	430
m	ft	0.3048
m^2	ft^2	9.29×10^{-2}
m^3	ft^3	2.83×10^{-2}
mg/Nm^3	gr/dscf	1.83
°C	°F	$t_c\,(^\circ C) = 5/9[t_f(^\circ F) - 32]$
KW	hp	0.746
joules/kg	Btu/lb	1.33×10^{-4}
m^3/sec/m^2	cfm/ft^2	5.08×10^{-3}
tonnes	tons (short)	0.907
kg	lb	0.454
m^3/sec	cfm	4.72×10^{-4}
m^3/sec	gpm	6.31×10^{-5}

OVERVIEW AND EVALUATION OF TWO YEARS OF OPERATION OF THE RIVERSIDE SPRAY DRYER SYSTEM

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ABSTRACT

Initial operation of the 100 MW spray dryer system at Northern States Power Company's Riverside Generating Station began in November 1980. At that time, a comprehensive test program was initiated to demonstrate the suitability of dry flue gas desulfurization (FGD) for utility coal fueled boilers. Since initial operation of the Riverside spray dryer began, other publications have described individual aspects of this system's performance. This paper provides a comprehensive analysis and overview of the performance test data collected during the initial two-year period of operation of the Riverside spray dryer.

Overall data correlations describing important variables in spray dryer operation and performance are established from test results obtained during operation of the Riverside spray dryer system under a wide range of conditions. Correlations between total and fabric filter sulfur dioxide (SO_2) removal and parameters such as lime stoichiometric ratio, total alkalinity, and approach temperature are presented for several different coals. Variations in moisture content of the solids collected in the spray dryer and fabric filter are evaluated to establish the sensitivity to a wide range of operating variables. In addition, system operation and control experiences are described to illustrate the interaction between flue gas flow, feed slurry flow, absorber outlet temperature, and SO_2 emissions during normal operation, as well as during transient conditions such as start-up, shutdown, and load swings. The effects of boiler soot blowing on the flue gas saturation temperature and system control are also discussed.

INTRODUCTION

Joy Manufacturing Company, Niro Atomizer Incorporated, and Northern States Power Company (NSP) jointly funded the 100 MW spray dryer demonstration program at the Riverside Generating Station. Design support for the spray dryer installation was provided by Black & Veatch, Engineers-Architects. The purpose of the Riverside demonstration program was to confirm that a full size spray dryer installed on a utility coal fueled power station could reliably and economically meet current SO_2 and particulate emission regulations. A single, full-size spray absorber module followed by a fabric filter was installed and began initial operation at Riverside Units 6 and 7 in November 1980. Testing of the Riverside system began in March 1981 and remains in progress.

This paper presents an overview of all performance test results obtained during 1981 and 1982. Overall data correlations are developed and relationships are established describing the sulfur dioxide removal efficiency and waste solids drying performance of the system. Fairly wide variations in some of the performance test results were observed over the extended test period. The observed range of performance appropriately reflects the actual performance of a full scale utility dry scrubber system subject to normal variations in operating conditions. Operational performance and control characteristics of the system are also presented. Major system parameters which are monitored and controlled during start-up, shutdown, and rapid load changes are discussed. The impacts of steam generator soot blowing on operation of the Riverside spray dryer are also described.

SYSTEM DESCRIPTION

Riverside Units 6 and 7 have identical Babcock & Wilcox steam generators which began operation in 1949 and 1950, respectively. The combined generating capacity of these units is approximately 100 MW. The design fuel for the steam generators is a 10,800 Btu/lb Illinois coal with a sulfur content of 3.5 per cent. Prior to the installation of the spray dryer and fabric filter system, each unit had individual electrostatic precipitators, ID fans, and chimneys.

The retrofit spray dryer system at Riverside treats the combined flue gas flow from Units 6 and 7. Figure 1 presents a simplified flow diagram of the spray dryer system. Flue gas at the air heater outlet from both units is combined and drawn through the single spray dryer module. Design features of the 46-foot diameter spray dryer module include a compound gas disperser and a single rotary atomizer. After flowing through the spray dryer, flue gas is directed to either the fabric filter or the original weighted wire precipitators. The fabric filter uses reverse gas cleaning and has twelve compartments. The gas flow at the discharge of the particulate collectors is divided and flows to the original Unit 6 and 7 chimneys.

Both a ball mill and an attrition slaker are installed for lime slurry preparation. Lime slurry is fed to a mix tank where it is mixed with recycled material from the spray dryer and particulate collector before being pumped to the atomizer. The recycle material is collected from the spray dryer and particulate collector (fabric filter or electrostatic precipitator) hoppers and consists of flyash, unreacted lime, and flue gas desulfurization reaction products. Material collected in the spray dryer and particulate collector hoppers which is not recycled is stored in silos prior to dry landfill disposal.

PERFORMANCE TEST RESULTS

During 1981 and 1982, 164 performance tests were conducted at Riverside. Most of the testing during this time consisted of short term parametric tests which were conducted to determine the influence of various individual parameters on system performance. Longer term system performance was evaluated by conducting five sets of demonstration tests. During the demonstration test periods, the system was maintained at relatively stable operating conditions for 5 to 10 days. Figure 2 summarizes the schedule for the performance tests which comprise the data presented herein.

During the two years of testing at Riverside, the operating conditions varied widely. Five different fuels were tested: Colstrip coal, Colstrip/Coke blend, Sarpy Creek coal, Sarpy Creek/Coke blend, and high sulfur Illinois coal. The Colstrip and Sarpy Creek coals are low sulfur coals from

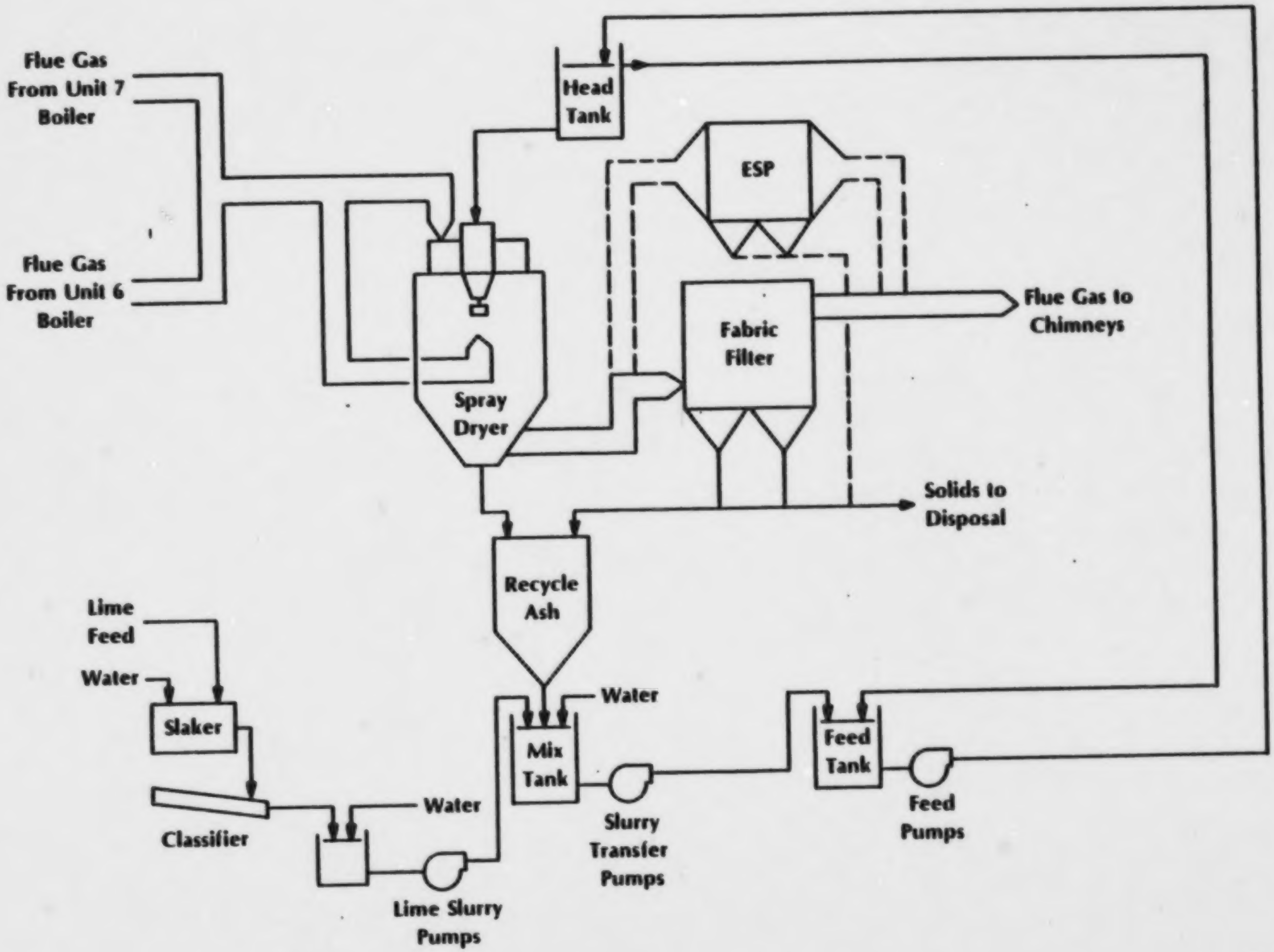


FIGURE 1 RIVERSIDE SPRAY DRYER SYSTEM

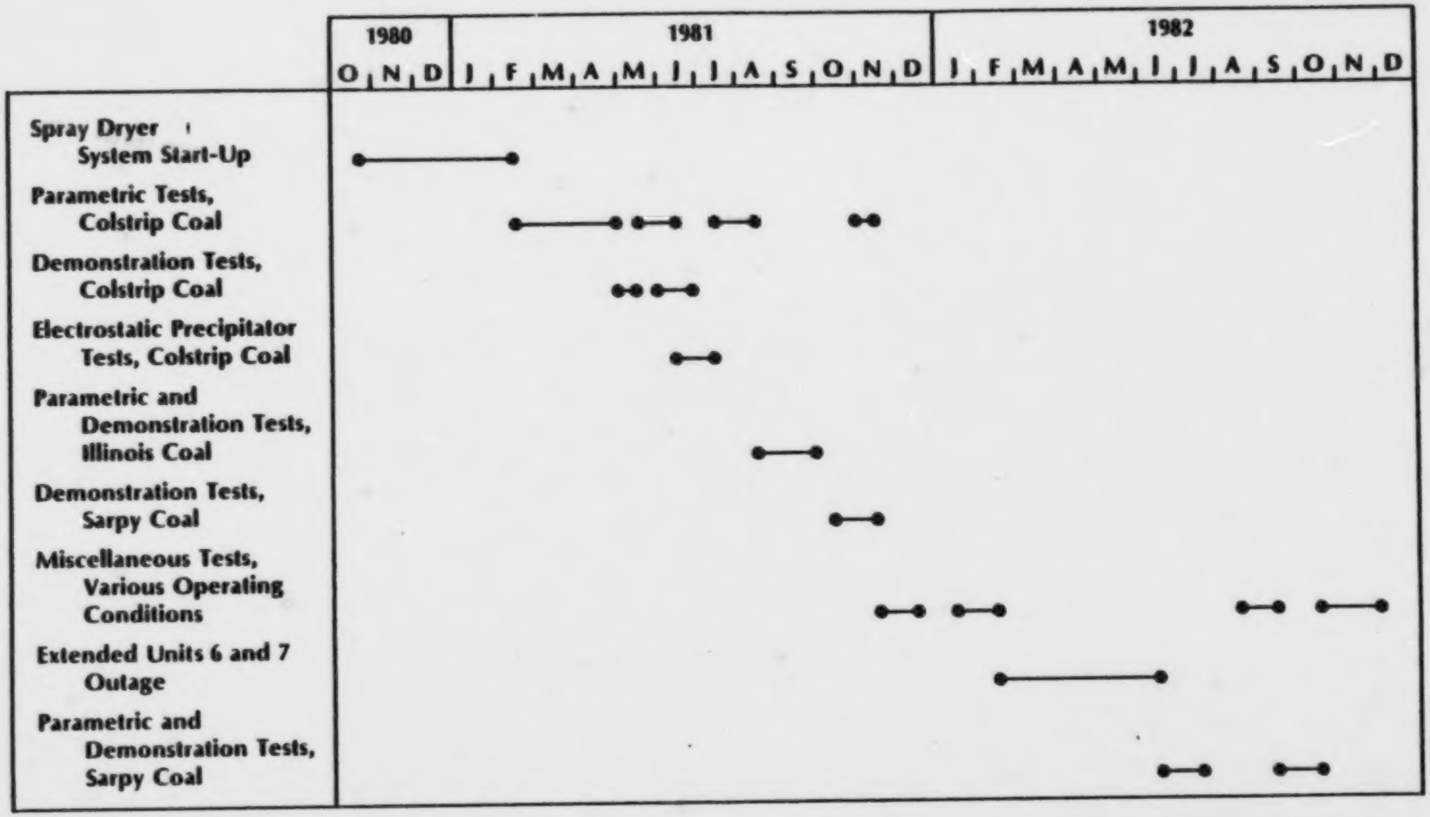


FIGURE 2 RIVERSIDE SPRAY DRYER PERFORMANCE TEST SCHEDULE

Montana. Table 1 presents some of the typical properties for each coal. Tests were performed with gas flows at the absorber inlet ranging from 34 to 114 per cent of design flow (170,000 acfm to 570,000 acfm). Approach temperatures ranged from 16 to 78 F. Feed slurry was atomized at fine, intermediate, coarse, and very coarse levels.

This paper focuses on the effects of a range of variable operating conditions on SO_2 removal performance and waste moisture content (dryness of the solid waste products). The test results are primarily viewed on the whole rather than as isolated individual groups of data or periods of testing to establish specific correlations. The results presented in this analysis should reflect the effects on performance of a wide range of operating conditions judged representative of those that full scale units will routinely experience.

SO_2 REMOVAL CORRELATIONS

LIME STOICHIOMETRIC RATIO

The ability to economically achieve SO_2 removal efficiencies required by New Source Performance Standards (NSPS) is a requirement for any flue gas desulfurization (FGD) system. As a measure of the Riverside dry scrubber performance, total SO_2 removal versus lime stoichiometric ratio (moles of lime added per mole of SO_2 removed) was plotted for all of the 164 performance tests. A multiple linear regression analysis was used to establish correlations between SO_2 removal, lime stoichiometric ratio, and approach temperature. Figure 3 presents the results of this analysis, applied to all 164 tests. The multiple linear regression technique of data correlation utilizes test data to generate a correlation between related parameters. Multiple linear regression is not a curve fitting technique. The curves presented on Figure 3 were generated using the regression equation established for total SO_2 removal and lime stoichiometric ratio. Curves were calculated for approach temperatures of 18 F and 40 F. These approach temperatures are typical of low and high approach temperatures frequently selected as design points for utility spray dryer installations. Since the tests were performed at a variety of approach temperatures, the distribution of test results around the regression curves is shown for comparative purposes only.

As shown on Figure 3, both the lime stoichiometric ratio and approach temperature affect the SO_2 removal. For a specific lime stoichiometric ratio, decreases in the approach temperature resulted in increased SO_2 removal. The form of the regression equation used to generate correlations between SO_2 removal and lime stoichiometry is as follows:

$$SRE = \frac{A}{LSR + AA} + B * \Delta T + K,$$

where
SRE = SO_2 removal efficiency, per cent
A, B = Regression coefficients
LSR = Lime stoichiometric ratio
AA = Ash Alkalinity constant
ΔT = Approach temperature, degrees F
K = Regression constant

Table 2 lists the coefficients and constants for the regression equation shown above, as well as the number of tests conducted using each fuel and the average SO_2 removal efficiency. The value

TABLE 1. TYPICAL PROPERTIES OF COALS AND FUEL BLENDS BURNED AT RIVERSIDE DURING 1981 AND 1982

Fuel	Heating Value lb/MBtu	Ash Content per cent	Sulfur Content per cent
Colstrip	8,700	8.3	0.8
Colstrip/Coke	9,200	6.7	1.2
Sarpy Creek	8,300	10.8	0.8
Sarpy Creek/Coke	9,200	8.9	1.2
Illinois	10,900	10.4	3.2

TABLE 2. REGRESSION EQUATION COEFFICIENTS FOR SO_2 REMOVAL AS A FUNCTION OF LIME STOICHIOMETRIC RATIO AND APPROACH TEMPERATURE FOR ALL RIVERSIDE TESTS

Fuel	Tests	Average SO_2 Removal %	Coefficients/Constants				Standard Error of Estimate %
			A	B	AA*	K	
Colstrip/Coke	94	85.9	-16.9	-0.65	0.42	114.9	9.1
Sarpy Creek/Coke	24	88.0	-79.3	-0.38	0.42	148.5	6.1
Illinois**	22	92.0		-0.36		99.9	3.9
Sarpy Creek	17	75.4	-29.2	-0.21	0.42	105.0	6.9
Sarpy Creek (no lime)	4	30.6					
Colstrip	3	69.3					
All Fuels	164	84.3	-33.8	-0.66	0.42	124.8	8.6

*The value for the ash alkalinity constant (AA) was established by optimizing the "all fuels" regression equation. The addition of such a constant to the other equations also improved the data correlations, and the same value was assumed for each case.

**For the limited range of data available, no significant correlation with lime stoichiometric ratio was found.

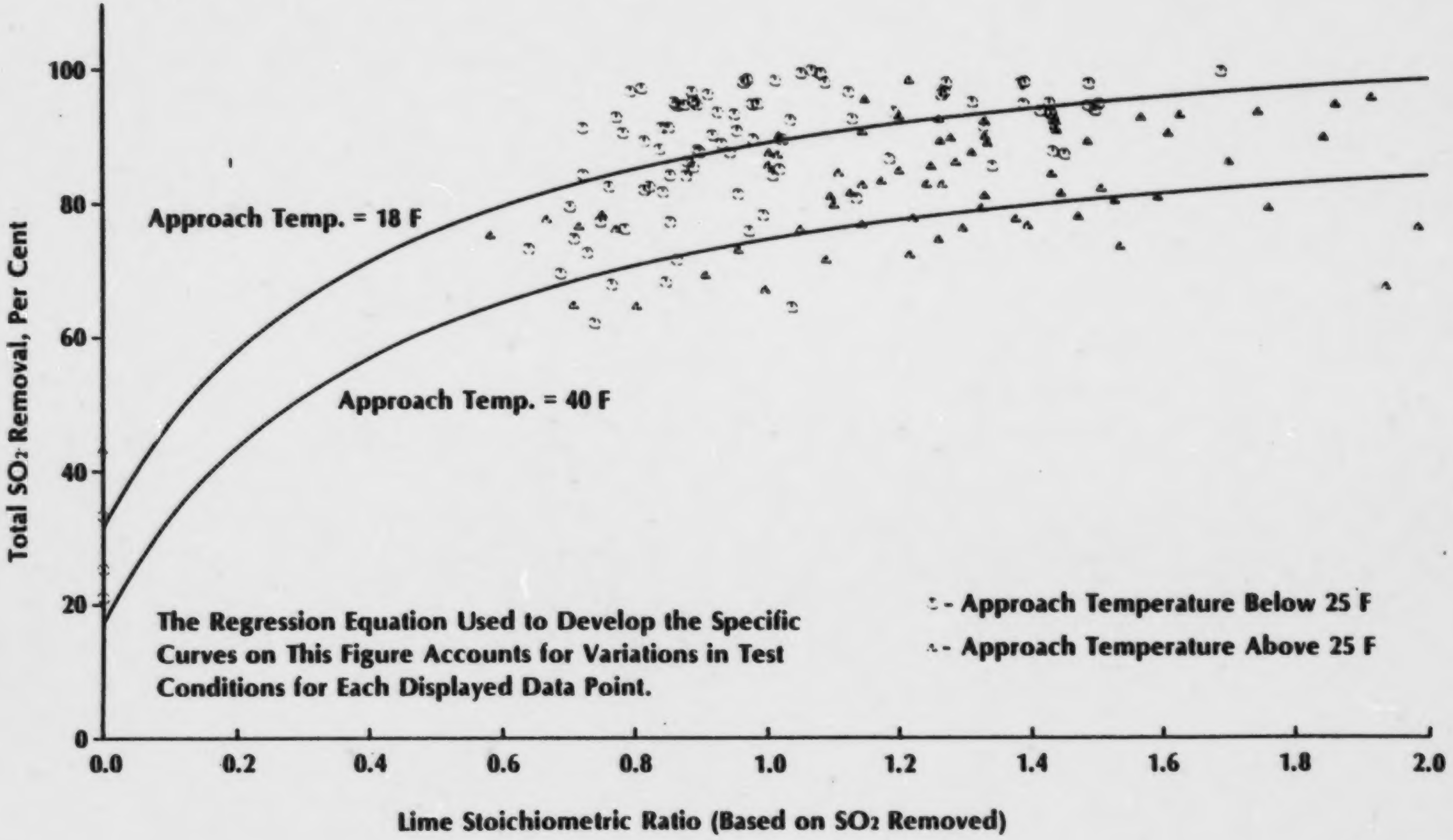


FIGURE 3 SO_2 REMOVAL AS A FUNCTION OF STOICHIOMETRIC RATIO AND APPROACH TEMPERATURE (ALL FUELS) of the ash alkalinity constant (AA) was selected by optimizing the regression equation using all of the test data. Inclusion of this same constant in the individual fuel regression equations improved these correlations as well. Insufficient data were available to optimize this value for each case.

Based on an analysis of the entire range of data, factors such as degree of atomization, gas flow, SO_2 inlet concentration, gas retention time, and amount of recycle could not be directly correlated to SO_2 removal for the limited variation in these values tested at Riverside. However, pilot testing has demonstrated that factors such as these are significant in spray dryer design and performance. The results show that for the full scale Riverside spray dryer system operating over a range of conditions, stoichiometric ratio, approach temperature, and ash alkalinity are the dominant process parameters which control SO_2 removal.

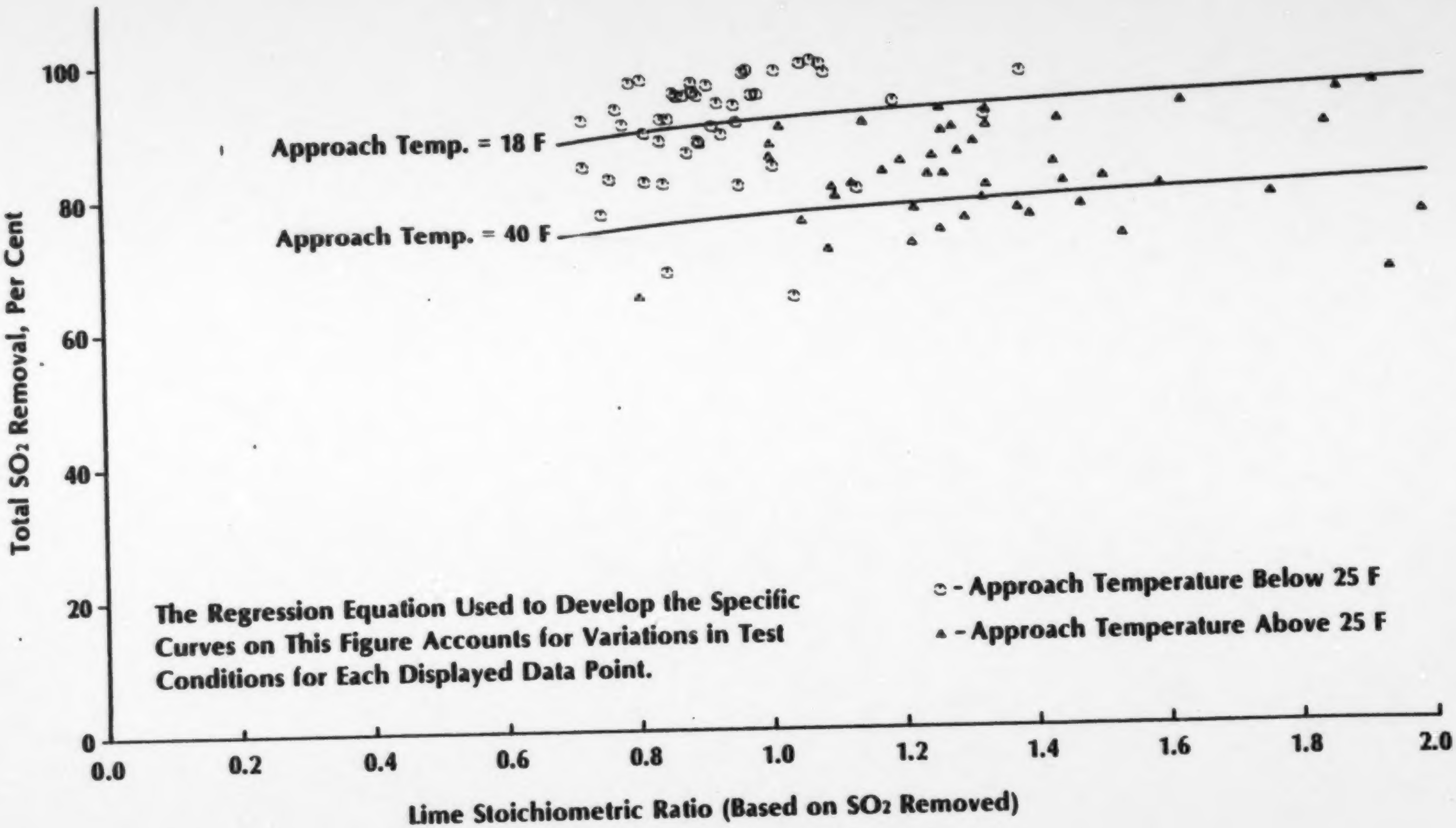
A separate regression analysis was performed on each fuel to determine individual contributions to the overall regression equation. Figures 4 through 6 show the SO_2 removal efficiency predicted by regression equations obtained for each fuel, plotted along with the respective data points. The Colstrip/Coke test data shown on Figure 4 correspond primarily with the mid-range of the SO_2 removal and lime stoichiometries for the overall correlation presented on Figure 3. The Sarpy Creek coal test data shown on Figure 5 generally correspond to the lower SO_2 removals and stoichiometries in the data base. The Illinois coal test data shown on Figure 6 comprise most of the high SO_2 removals and corresponding high lime stoichiometries in the data base. The regression equations obtained from the analysis of test results from each fuel, indicate varying sensitivities to approach temperature and lime stoichiometric ratio. The Sarpy Creek coal is the most sensitive to lime stoichiometry but the least sensitive to apporach temperature. The Illinois coal did not exhibit much sensitivity to stoichiometry but was sensitive to approach temperature. The Colstrip/Coke blend was the most sensitive to approach temperature and it also was moderately sensitive to stoichiometric ratio.

TOTAL ALKALINITY RATIO

The alkalinity ratio is a measure of the total alkaline material which is fed to the atomizer. In addition to the fresh lime, the alkaline material includes alkalinity in the ash and unreacted lime in the recycled solids. Since the lime stoichiometric ratio provided a good correlation with total SO_2 removal, the alkalinity ratio was expected to provide a better correlation. Total SO_2 removal and the corresponding alkalinity ratio for all 164 tests are shown on Figure 7. The overall trend of data on this figure indicates that there is a correlation between SO_2 removal and alkalinity ratio. However, a significant number of tests have alkalinity ratios of less than 1.0. Since alkalinity ratio is defined in this case as equivalent moles of CaO per mole of SO_2 removed, the minimum theoretical alkalinity ratio is 1.0. During the Riverside test program two different techniques were used to measure the total alkalinity ratio. The results obtained using each technique differed and neither technique appeared to provide consistent and reasonable results. Thus, no correlations were made using this parameter.

SO_2 REMOVAL IN THE PARTICULATE COLLECTOR

The total SO_2 removal which occurs across a dry scrubber system consists of SO_2 removed in the spray dryer and SO_2 removed in the particulate collector. SO_2 removal in the particulate collector results from the additional contact of the alkaline solids with the remaining SO_2 in the flue gas. SO_2 removal in the particulate collector can be affected by process parameters in a different



The Regression Equation Used to Develop the Specific Curves on This Figure Accounts for Variations in Test Conditions for Each Displayed Data Point.

FIGURE 4 SO_2 REMOVAL AS A FUNCTION OF STOICHIOMETRIC RATIO AND APPROACH TEMPERATURE (COLSTRIP/COKE FUEL)

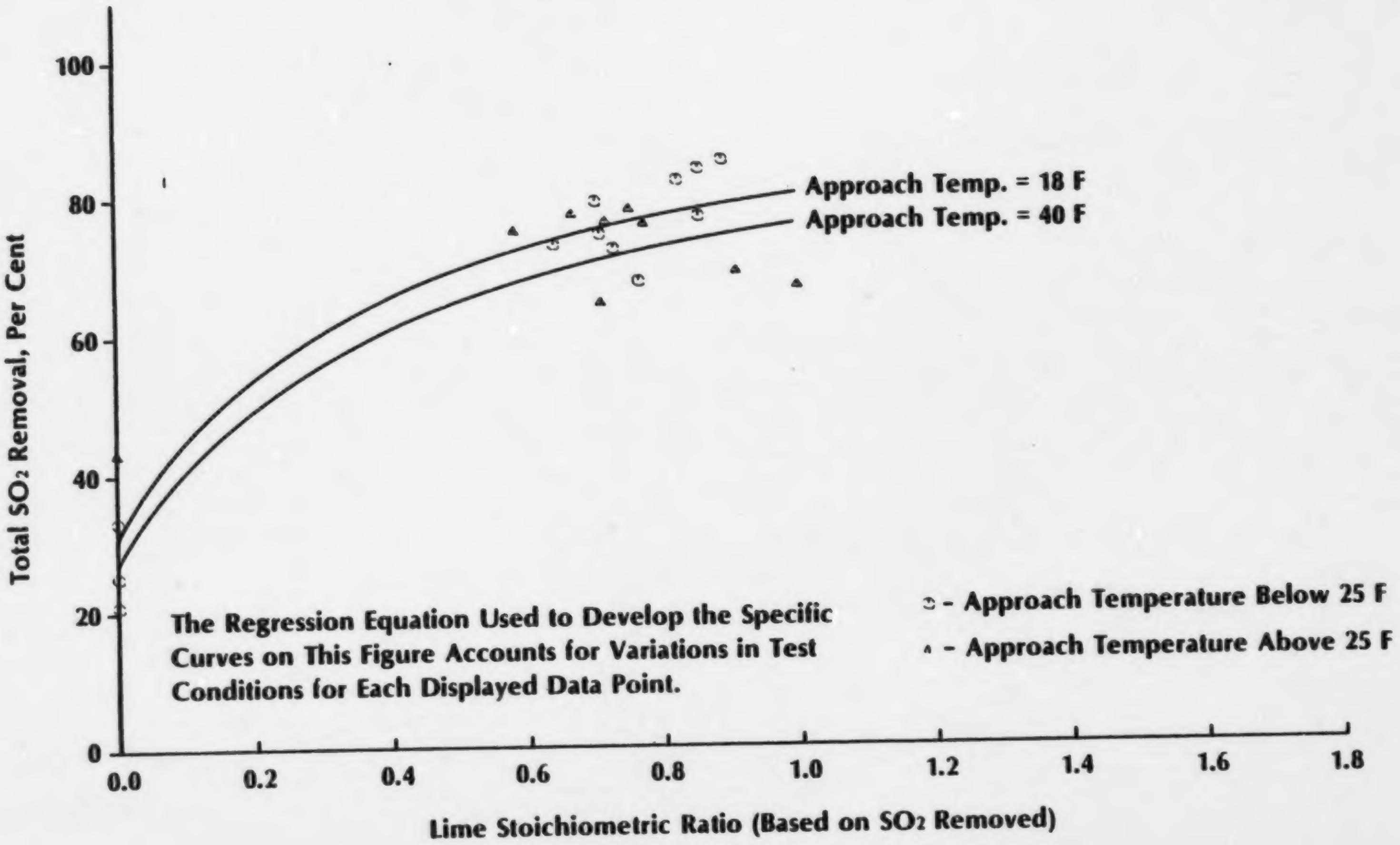


FIGURE 5 SO_2 REMOVAL AS A FUNCTION OF STOICHIOMETRIC RATIO AND APPROACH TEMPERATURE (SARPY CREEK COAL)

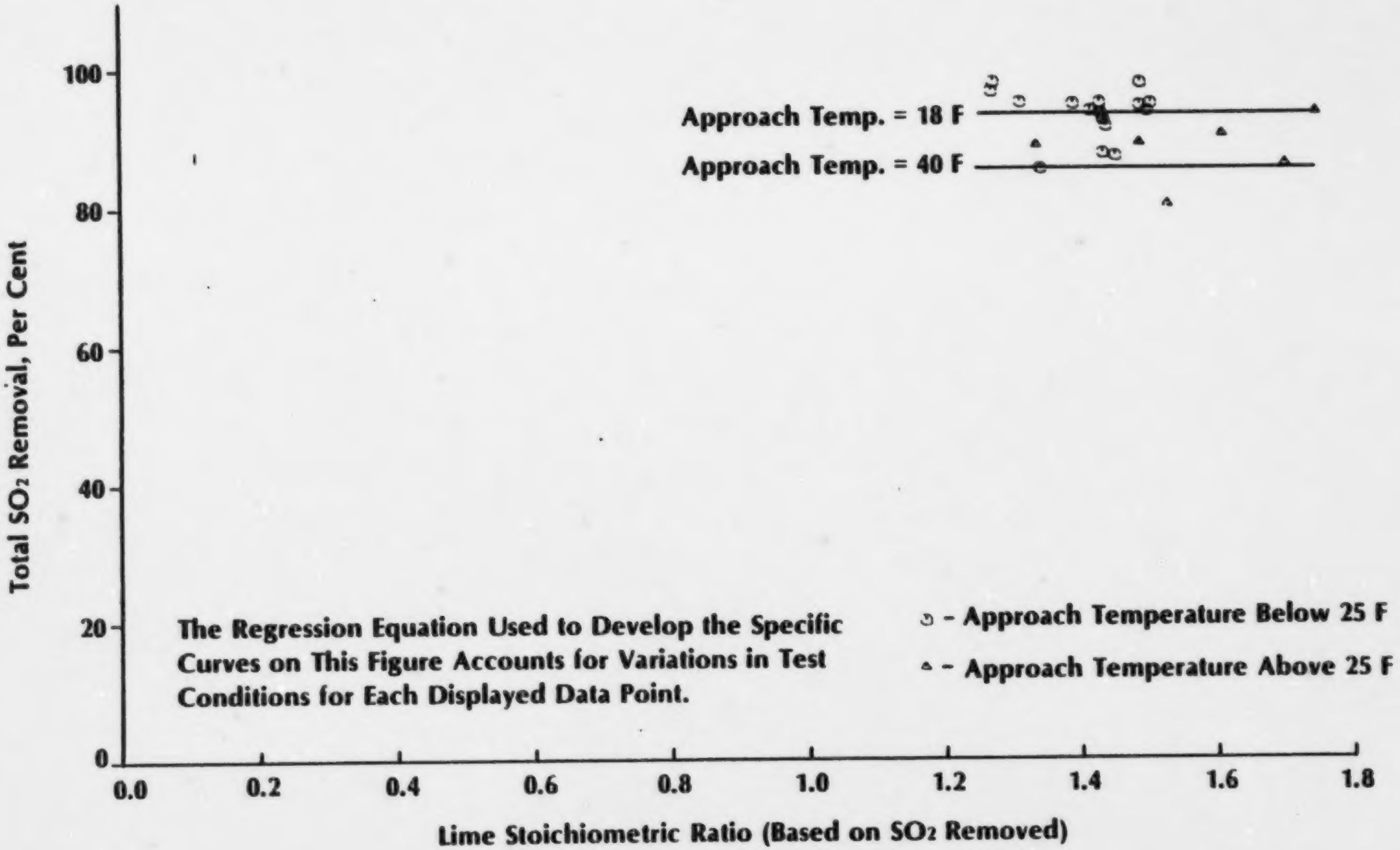


FIGURE 6 SO₂ REMOVAL AS A FUNCTION OF STOICHIOMETRIC RATIO AND APPROACH TEMPERATURE (ILLINOIS COAL)

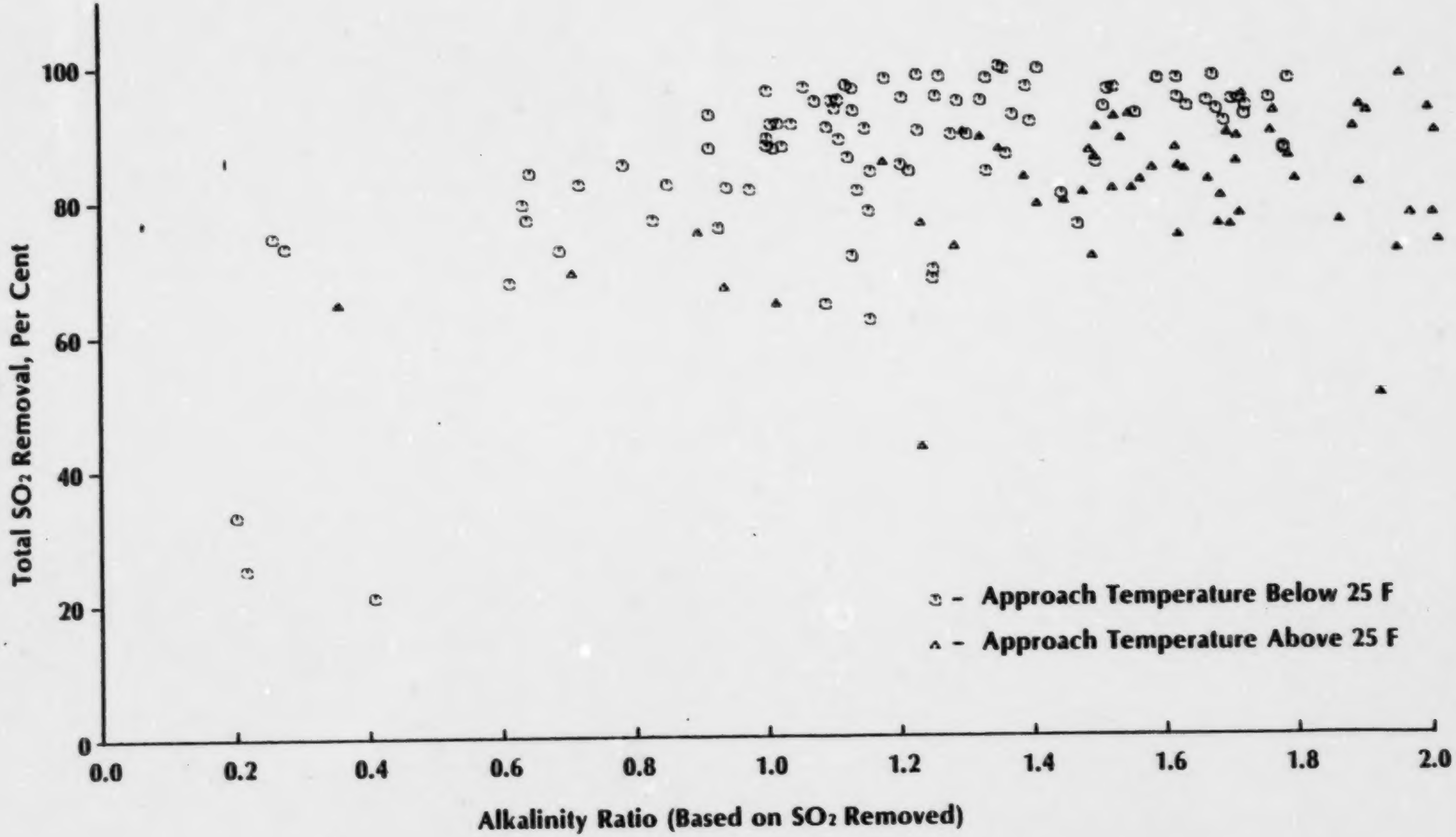


FIGURE 7 SO_2 REMOVAL AS A FUNCTION OF ALKALINITY RATIO AND APPROACH TEMPERATURE (ALL FUELS) manner than SO_2 removal in the spray dryer. To determine the sensitivity of SO_2 removal in the particulate collector to various process parameters, a separate analysis was performed.

Figure 8 presents total and fabric filter SO_2 removal efficiency as a function of lime stoichiometric ratio. As shown in this figure, fabric filter SO_2 removal efficiencies generally range between 10 to 20 per cent and average approximately 15 per cent. Figure 9 presents total and electrostatic precipitator SO_2 removal as a function of lime stoichiometric ratio. As shown in this figure, electrostatic precipitator (ESP) SO_2 removal ranges from 0 to 20 per cent and averages approximately 6 per cent. Higher average SO_2 removal efficiencies in the fabric filter than in the ESP are attributed to more effective interaction between the flue gas and the particulate in the filter cake on the filter bag surfaces. Although measurements were not consistently available from the performance test data, factors such as fabric filter gas-to-cloth ratio, pressure drop, and frequency of bag cleaning are expected to have a significant effect on fabric filter SO_2 removal.

An analysis was performed to determine the sensitivity of fabric filter SO_2 removal to the inlet SO_2 concentration. Figure 10 presents a comparison of these parameters for all the fuels. No significant correlation could be found in the Riverside data between fabric filter SO_2 removal and inlet SO_2 concentration. A similar analysis performed on each fuel individually did not yield an improved correlation. Figure 11 presents a comparison of the Riverside fabric filter SO_2 removal as a function of approach temperature for all the fuels. Flue gas approach temperature to the dew point establishes the relative moisture content of the flue gas. High flue gas moisture content should result in improved SO_2 removal in the fabric filter. Contrary to what was expected, there appeared to be no correlation between fabric filter SO_2 removal and approach temperature. A similar analysis was performed on each of the fuels; however, no significant correlations were obtained. Correlations between fabric filter SO_2 removal and approach temperature have been shown to exist in laboratory and pilot scale units. However, on a full scale operating unit, variations in other process parameters such as fabric filter gas-to-cloth ratio and frequency of cleaning may diminish the influence of approach temperature on SO_2 removal.

An interesting correlation which was apparent from the test data was the correlation between fabric filter SO_2 removal and spray down temperature. Spray down temperature is the difference between the spray dryer inlet and outlet temperatures. Spray down temperature determines the total moisture content of the flue gas but is independent of the dew point temperature. Figure 12 shows the regression equation curve established for these parameters. As shown on Figure 12, fabric filter SO_2 removal appears to increase with increased spray down temperatures. The composite regression equation developed for all fuels (all tests with the fabric filter) for fabric filter SO_2 removal and spray down temperature is as follows:

$$\text{FFSRE} = \frac{A}{\text{SDT}} + K,$$

where
FFSRE = Fabric filter SO_2 removal efficiency, per cent
A = Regression coefficient (-2204.6)
SDT = Spray down temperature, per cent
K = Regression constant (28.2)

Analysis of fabric filter SO_2 removal versus spray down temperature for each individual fuel did not produce improved correlations. The Riverside spray dryer is operated to maintain a fixed

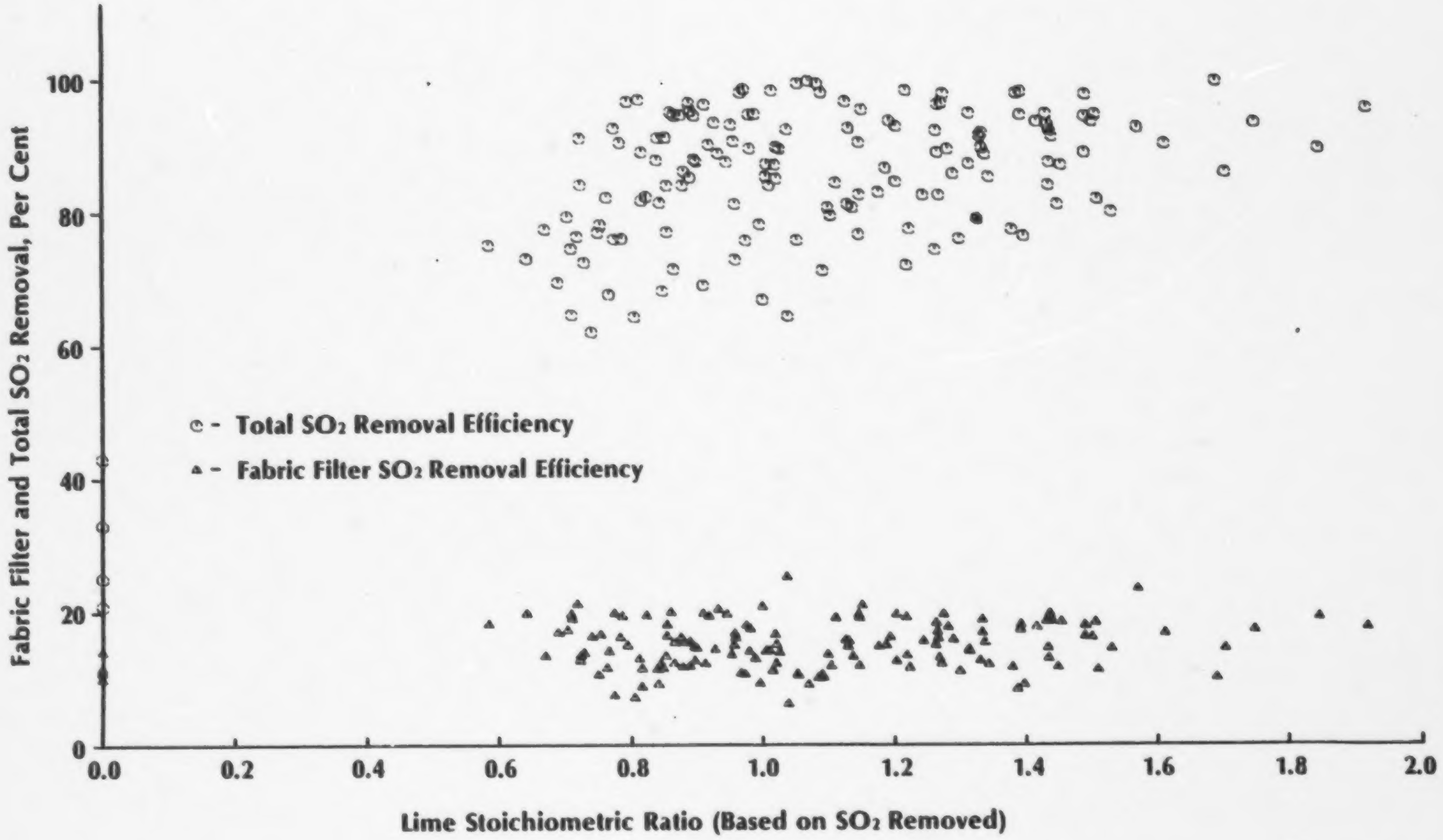


FIGURE 8 TOTAL AND FABRIC FILTER SO_2 REMOVAL AS A FUNCTION OF LIME STOICHIOMETRIC RATIO (ALL FUELS)

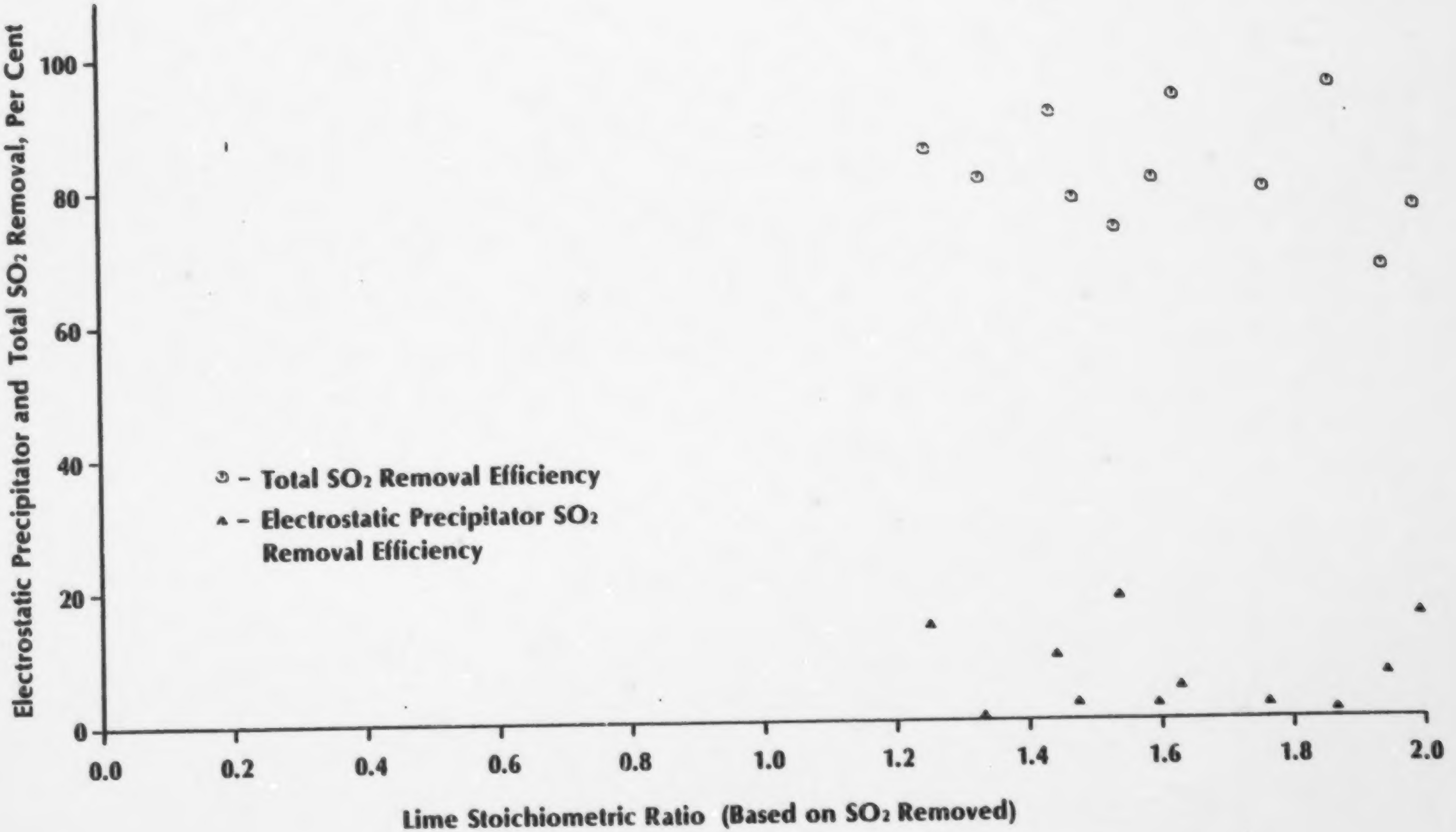


FIGURE 9 TOTAL AND ELECTROSTATIC PRECIPITATOR SO_2 REMOVAL AS A FUNCTION OF LIME STOICHIOMETRIC RATIO (COLSTRIP/COKE FUEL)

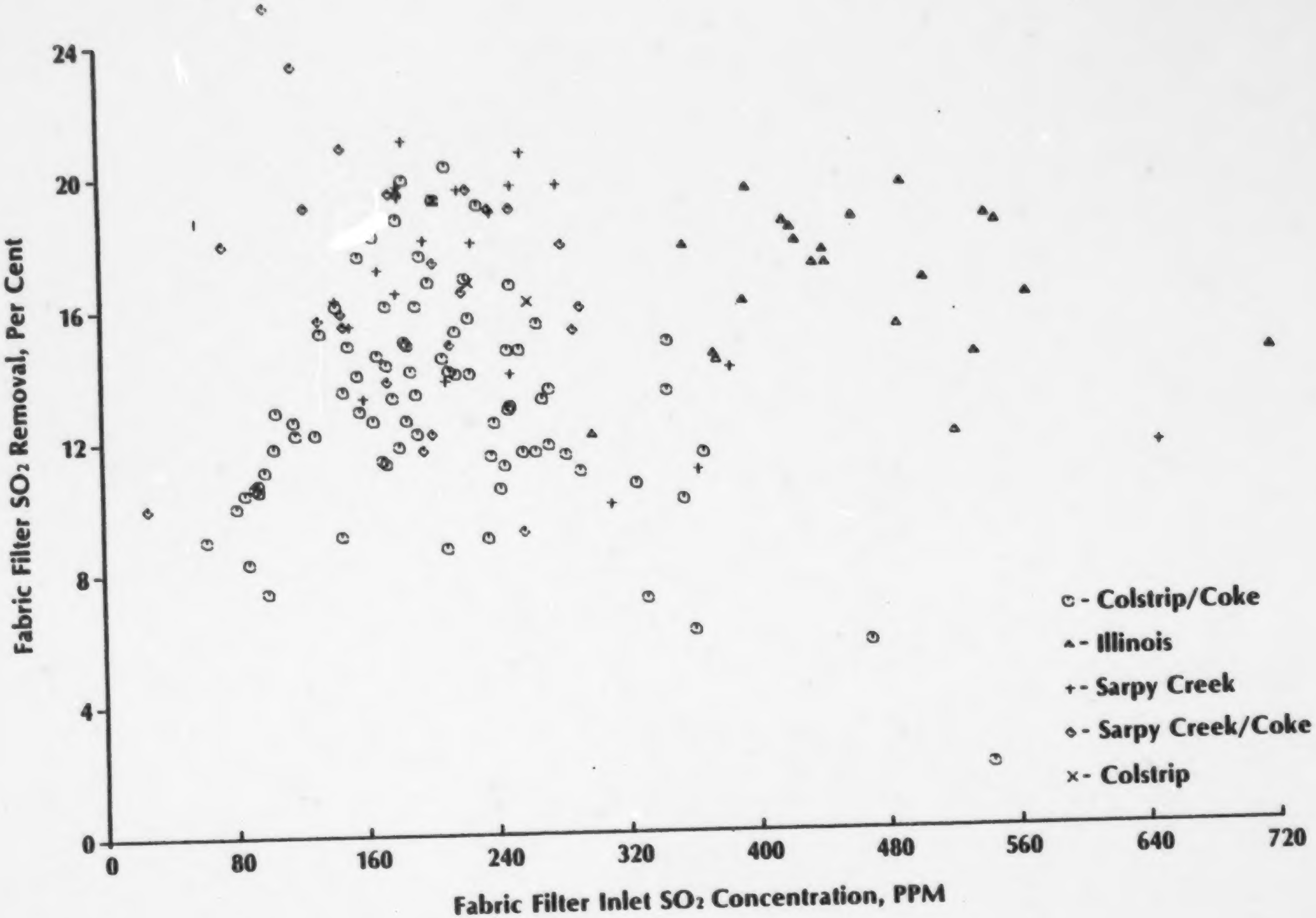


FIGURE 10 FABRIC FILTER SO_2 REMOVAL AS A FUNCTION OF FABRIC FILTER INLET SO_2

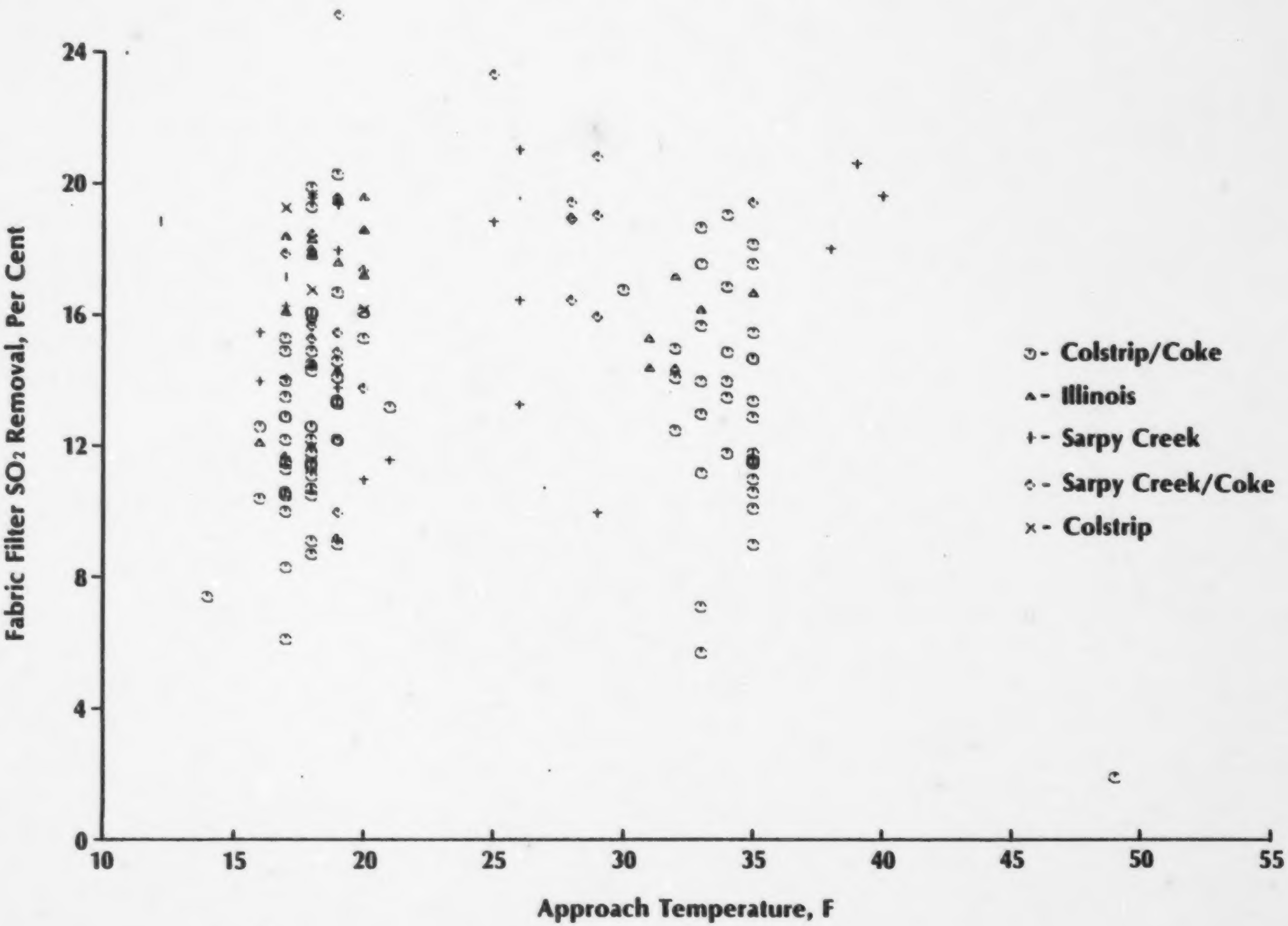


FIGURE 11 FABRIC FILTER SO_2 REMOVAL AS A FUNCTION OF APPROACH TEMPERATURE

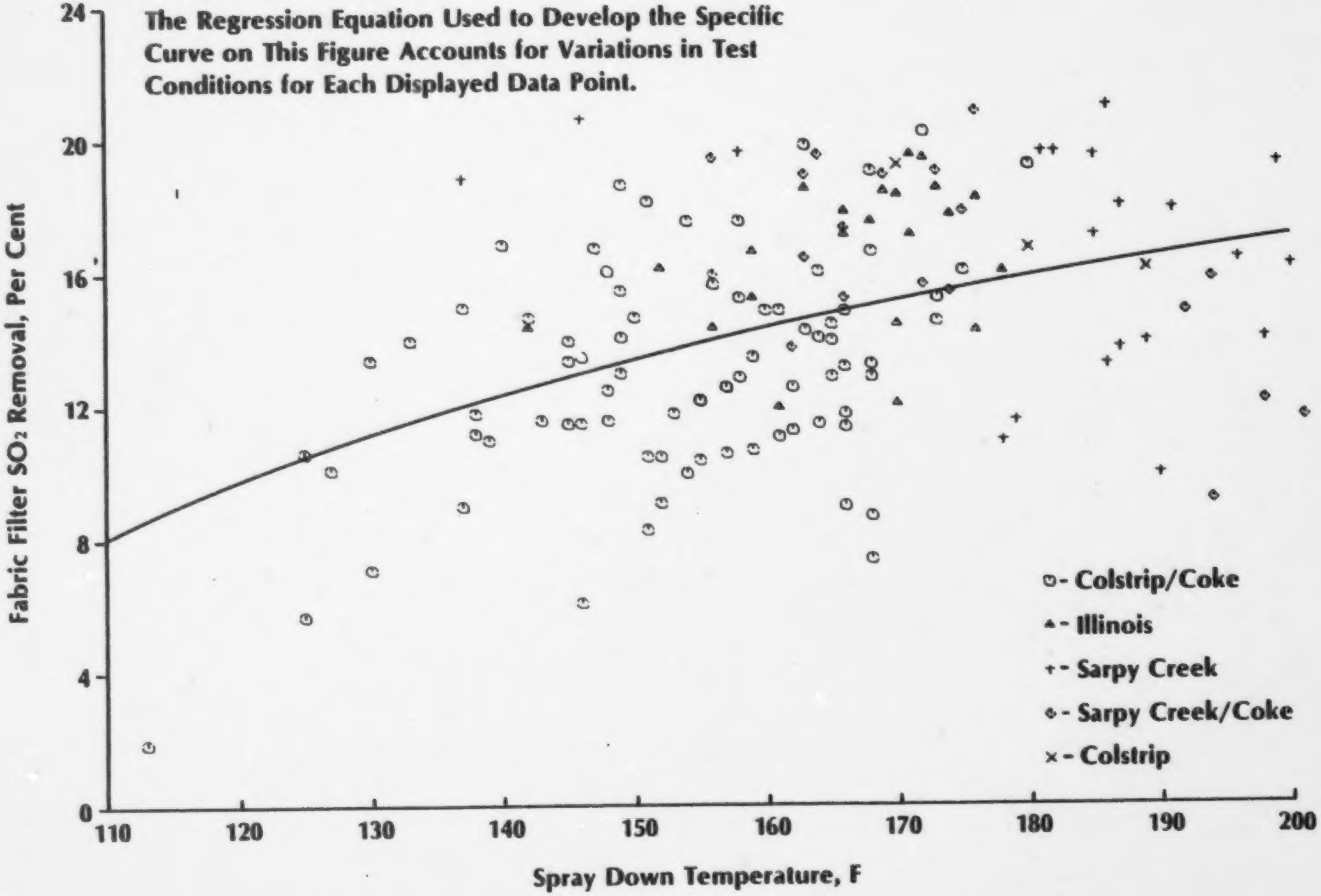


FIGURE 12 FABRIC FILTER SO_2 REMOVAL AS A FUNCTION OF SPRAY DOWN TEMPERATURE

proportion of solids and water in the atomizer feed. Thus, increasing the spray down temperature also increases the amount of particulate matter in the flue gas at the fabric filter inlet. From the Riverside data it appears that the greater the amount of particulate matter entering the fabric filter, the greater the amount of SO_2 removed in the fabric filter. Higher particulate concentrations in the fabric filter may provide a higher ratio of alkaline material on the bags which improves SO_2 removal efficiency. Other parameters such as gas-to-cloth ratio, pressure drop, and cleaning frequency may also contribute to this correlation.

Due to the small size of the data base, no attempts were made to correlate precipitator SO_2 removal with any of the process parameters. As previously discussed, the average SO_2 removal was less in the ESP than in the fabric filter. However, the Riverside precipitators are relatively small and were not designed for spray dryer service. New dry scrubber installations using precipitators may achieve slightly higher SO_2 removal efficiencies in the particulate collector than those achieved using the precipitators at Riverside.

WASTE MOISTURE CORRELATIONS

The ability of the spray dryer to produce a waste product which is relatively dry is a critical requirement of the system. As a measure of the ability of the Riverside dry scrubber to achieve a dry product, the moisture content of the solids discharged from the spray dryer cone and fabric filter hoppers was periodically measured during each performance test.

The average and peak moisture contents of the solids recorded during the performance tests are as follows.

	Average Moisture Content (per cent)	Peak Moisture Content (per cent)
Spray Dryer Solids	1.5	4.0
Fabric Filter Solids	1.0	2.3

Generally, the solids were sufficiently dry to be conveyed by conventional ash handling equipment.

To determine the effect of major process parameters on the waste moisture content, a comparison of spray dryer waste moisture and approach temperature was made. Figure 13 shows the corresponding approach temperature for each waste moisture content. Although there is a general trend for decreasing waste moisture with increasing approach temperature, a strong correlation was not apparent from the Riverside test data. Since approach temperature is a relative indication of the amount of moisture in the flue gas, it was expected to correlate with spray dryer waste moisture. Additional comparisons of waste moisture with spray down temperature, feed solids content, atomization, and mass flow rate were made. In each case, no strong correlation with waste moisture was apparent. A similar analysis was performed on each fuel, but no significant correlations were found. The range of recorded spray dryer waste moisture at Riverside is very narrow. Correlations may exist between these parameters; however, on a full scale operating unit, the range of operating conditions would not normally fluctuate enough for these correlations to be apparent.

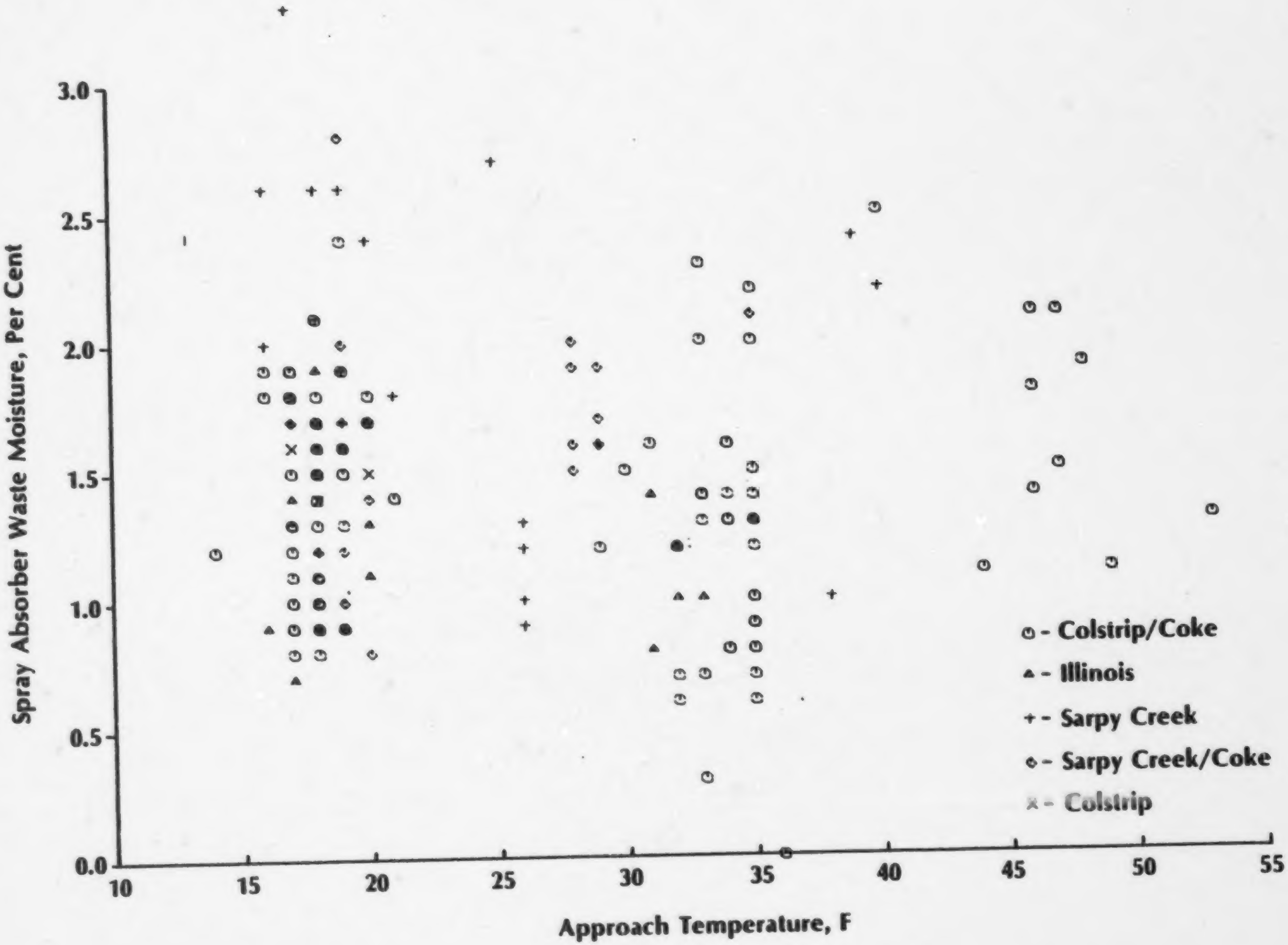


FIGURE 13 SPRAY ABSORBER WASTE MOISTURE AS A FUNCTION OF APPROACH TEMPERATURE

One correlation which was apparent from the test data was the correlation between spray dryer and fabric filter waste moisture. Figure 14 shows this correlation. As presented on Figure 14, fabric filter waste moisture content increases linearly with spray dryer waste moisture. The composite equation developed for all fuels (all tests) is as follows:

$$FFWM = A * SDWM,$$

where
FFWM = Fabric filter waste moisture, per cent
A = Regression coefficient (0.67)
SDWM = Spray dryer waste moisture, per cent

This correlation reflects the longer drying time of the solids in the fabric filter. Flue gas passing through the filter cake on the fabric filter bags provides additional drying of the collected solids.

OPERATIONAL PERFORMANCE AND CONTROL

SYSTEM CONTROL

A discussion of the primary control parameters is necessary before discussing the effects of start-up, shutdown, and rapid load swings on the performance of the spray dryer. Control of the Riverside spray dryer is based on maintaining a constant approach temperature at the spray dryer outlet. The approach temperature is held constant by maintaining the required evaporation rate in the spray dryer to quench the flue gas to the desired temperature above the adiabatic saturation temperature. Moisture is added to the gas stream through the atomizer with the additive feed slurry, while the fraction of solids in the atomizer feed is maintained at a constant level of approximately 35 per cent. To decrease the approach temperature, the total flow to the atomizer is increased.

To control the SO_2 removal efficiency of the dry scrubber, the ratio of lime and recycled solids in the feed slurry is changed, but the total amount of solids and water in the atomizer feed remains constant. To increase SO_2 removal, the ratio of the lime to feed solids is increased and the amount of recycle material is decreased. Conversely, to decrease SO_2 removal, the amount of lime in the atomizer feed is decreased and the amount of recycle material is increased. Due to the residence time in the slurry preparation system, changes in SO_2 removal efficiency can not be made instantaneously unless a change in approach temperature is also desired. Since the approach temperature has a significant effect on SO_2 removal, it is difficult to stabilize the absorber outlet temperature if approach temperature is used to control SO_2 removal. However, since only an average SO_2 removal rate is required, periods of low SO_2 can be offset by periods of higher SO_2 removal.

SYSTEM START-UP

The start-up of the Riverside spray dryer can be accomplished whenever the flue gas temperature and flue gas flow rate are high enough to allow proper drying in the chamber. At Riverside, a minimum gas flow of 125,000 ft^3/min (25 per cent gas flow) at 200 F is necessary to meet the minimum requirements for spray dryer operation. Start-up of the spray dryer is performed by an automatic control system which requires only limited operator interface once feed slurry flow is

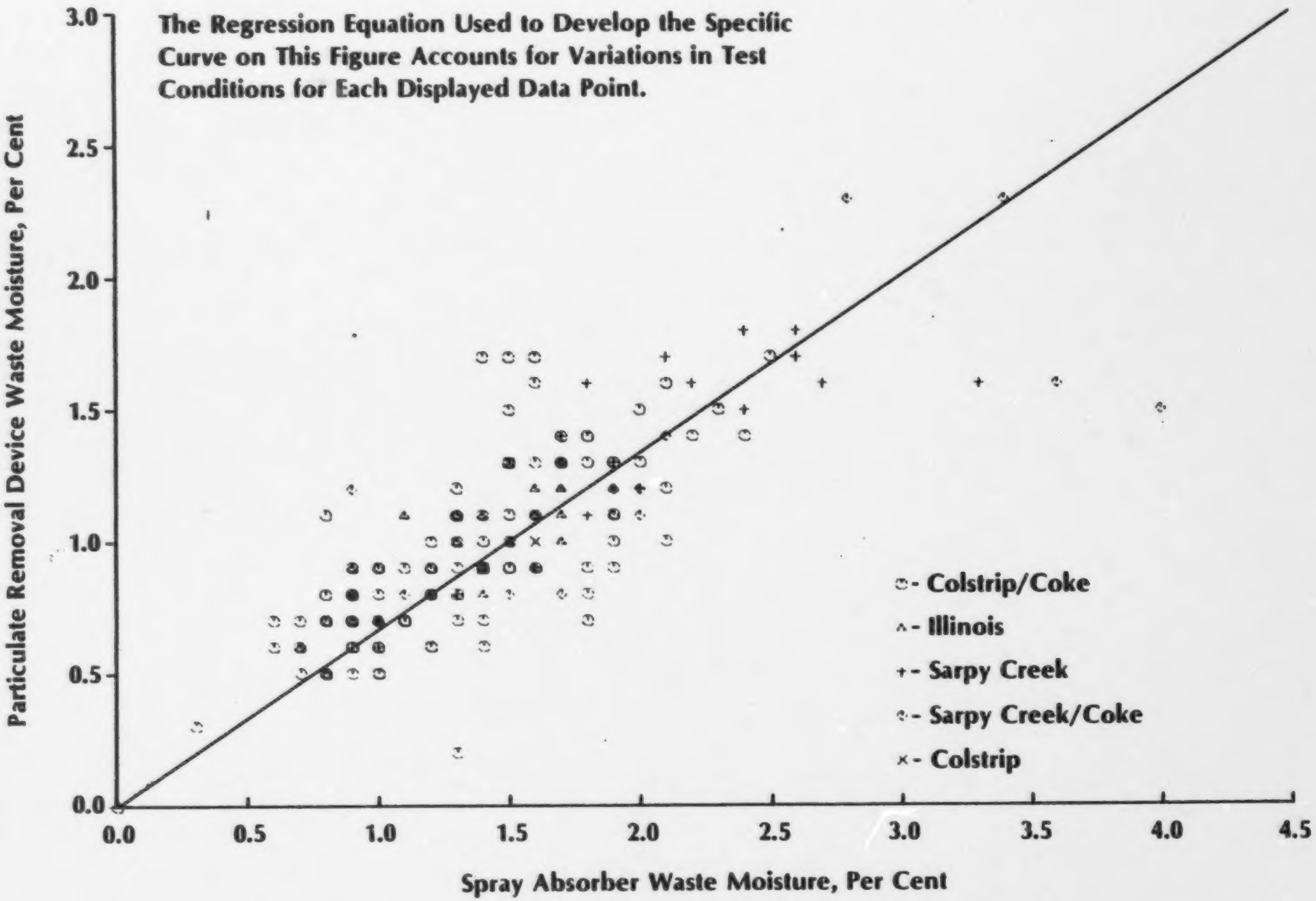


FIGURE 14 PARTICULATE REMOVAL DEVICE WASTE MOISTURE AS A FUNCTION OF SPRAY ABSORBER WASTE MOSITURE

established to the atomizer. Figure 15 shows a typical warm start-up of the spray dryer at Riverside. As shown on Figure 15, the gas flow is increased as the boilers come on line, and SO_2 is generated when coal firing begins. The feed slurry flow to the atomizer is initiated when the minimum gas flow and temperature are reached. SO_2 removal begins immediately when feed slurry flow is established. As the boilers continue to gain load, the slurry flow to the atomizer increases accordingly to maintain the desired absorber outlet temperature. When the absorber outlet temperature has stabilized, the automatic SO_2 removal control system adjusts the amount of lime in the feed to achieve the desired SO_2 removal. The time required for complete stabilization at the desired SO_2 removal is dependent primarily on the composition of the feed stock prior to start-up. The time to reach stable SO_2 removal also varies for hot, warm, or cold steam generator start-ups, because of differing initial flue gas temperatures.

SYSTEM SHUTDOWN

The flue gas flow and temperature requirements used for the shutdown of the spray dryer are the same as for the start-up, but it is easier to reduce transient SO_2 peaks during a controlled boiler shutdown because sufficient gas flow and temperature are present to allow spray drying until after the steam generator fire is extinguished. The feed slurry to the atomizer is stopped when the gas flow is no longer sufficient to ensure proper drying and the atomizer is flushed and taken out of service. As shown in Figure 16, the spray dryer follows load as the boilers are ramping down and comes off line without SO_2 emissions increasing.

RAPID LOAD SWINGS

The Riverside spray dryer is often subject to rapid load swings. As shown in Figure 17, the system automatically responds to the increased and decreased gas flow. The feed slurry flow rate follows closely to the gas flow, resulting in an almost constant outlet temperature. The load change illustrated on this figure was accomplished by bringing one of the boilers on line after the other was operating at steady load. This results in some dilution of the SO_2 concentration into the spray dryer and accounts for the reduced emissions during this period.

WET BULB TEMPERATURE FLUCTUATIONS

To control a spray dryer at a constant approach temperature, it is necessary to measure the adiabatic saturation (wet bulb) temperature of the flue gas. At Riverside, wick-and-water reservoir devices are installed at the fabric filter outlet to measure and transmit wet bulb temperature to the control system. It is necessary to locate these devices downstream of the fabric filter to minimize the fouling of the wick by particulate matter. Although the wet bulb measurement devices provide an accurate and reasonably reliable indication of the wet bulb temperature downstream of the baghouse, they do not provide precise information on the wet bulb temperature in the spray dryer chamber. Fluctuations in wet bulb temperature at the spray dryer inlet are not transmitted to the control system until 30 to 90 seconds after they occur. In addition, by the time the gas passes through the fabric filter and reaches the outlet duct, the heat transfer and air leakage reduce the wet bulb temperature by 1 to 3 F from that at the spray dryer inlet.

To determine the type of wet bulb fluctuations which may occur at the spray dryer inlet, a temporary device was installed and continuously monitored. Figure 18 presents the wet bulb temperature fluctuations at the inlet of the spray dryer during a typical soot blowing sequence using

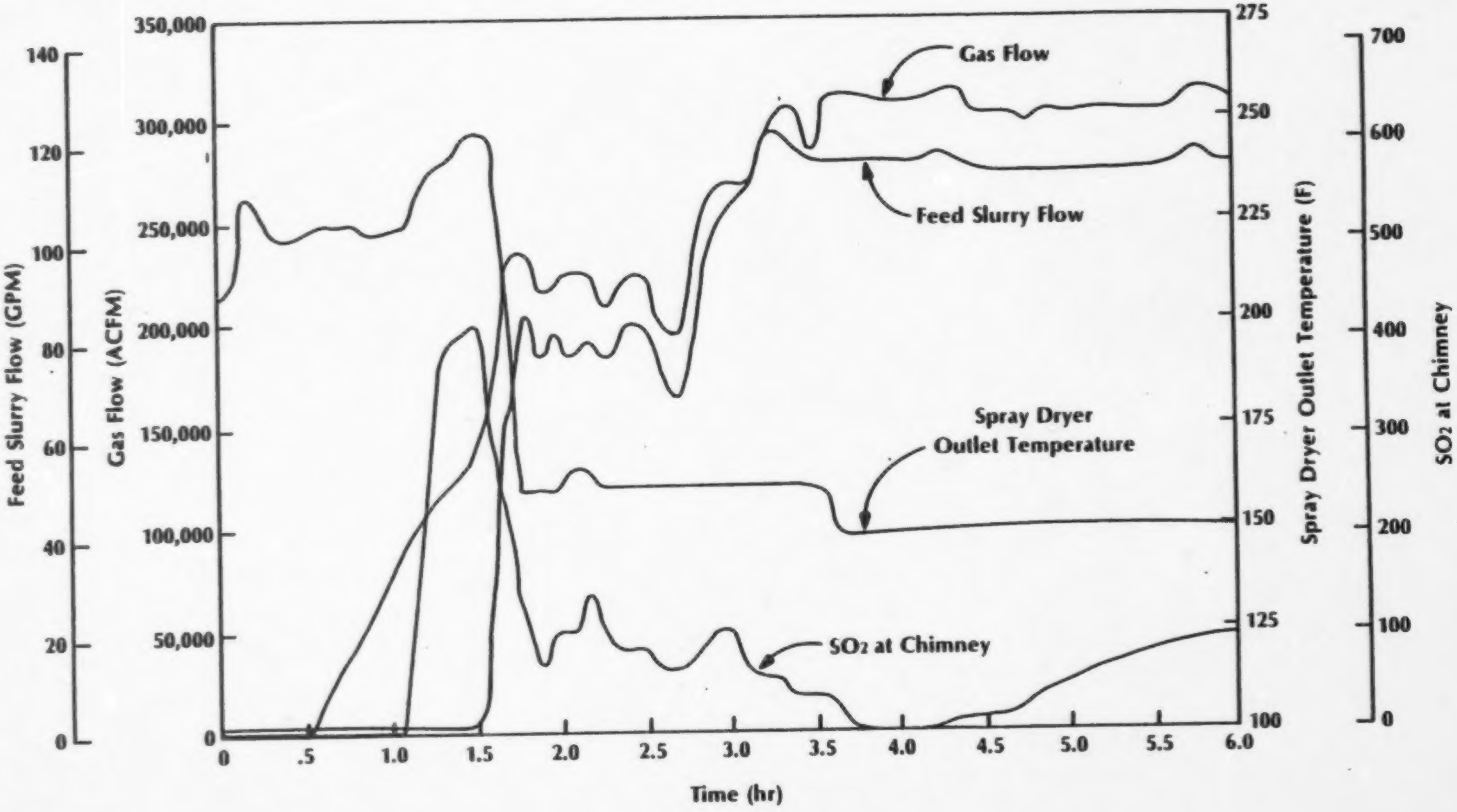


FIGURE 15 RIVERSIDE SPRAY DRYER SYSTEM RESPONSE DURING START-UP

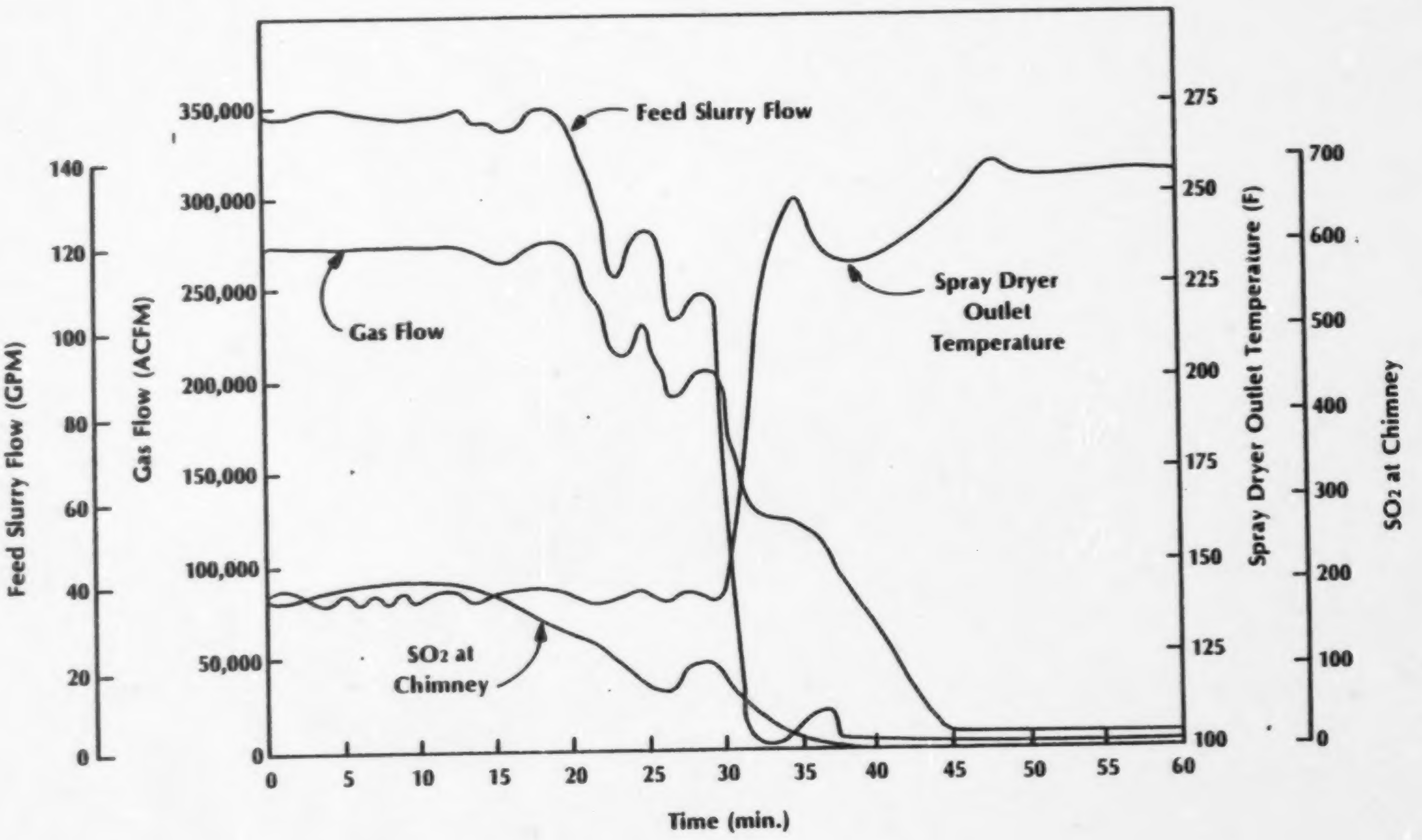


FIGURE 16 RIVERSIDE SPRAY DRYER RESPONSE DURING SHUTDOWN

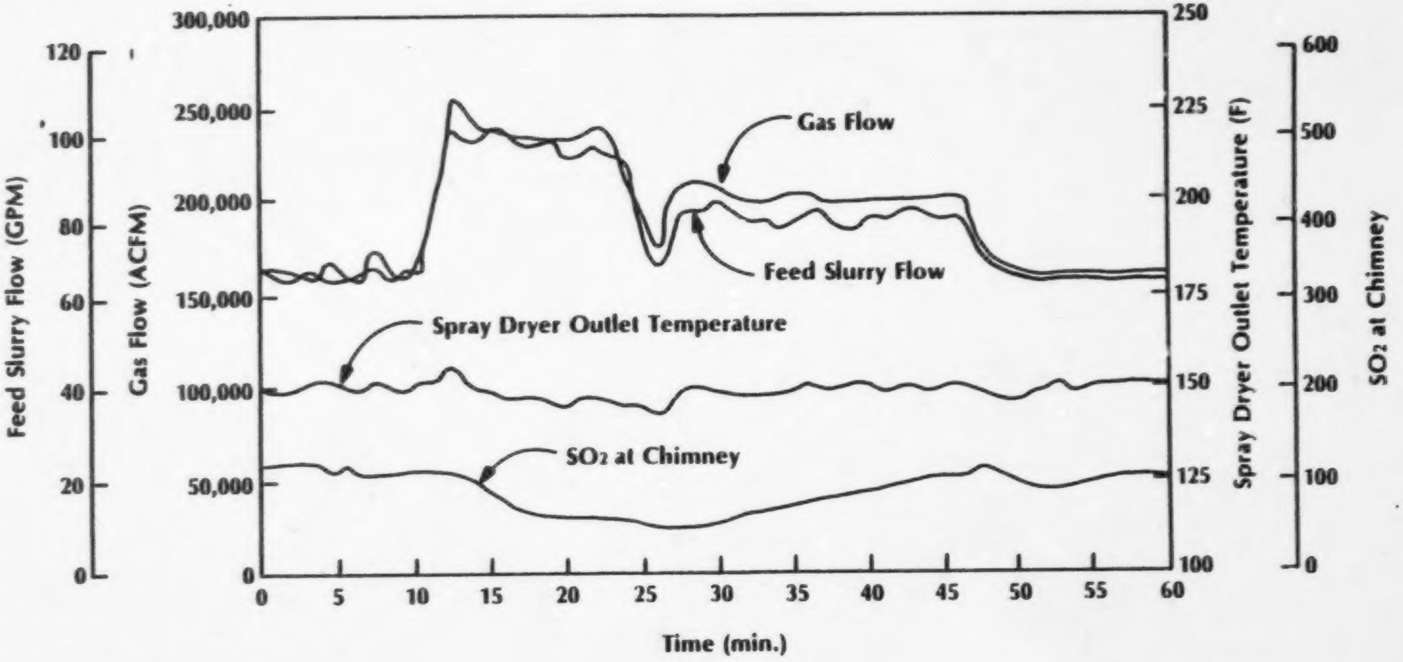
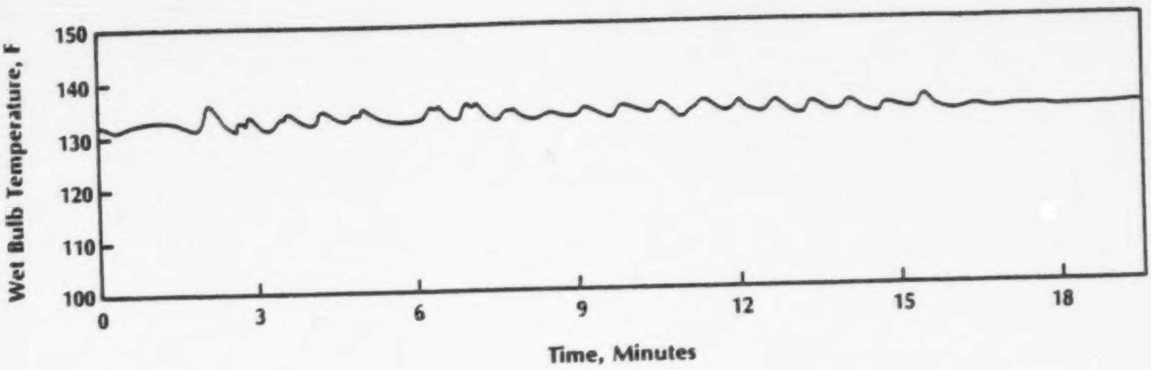
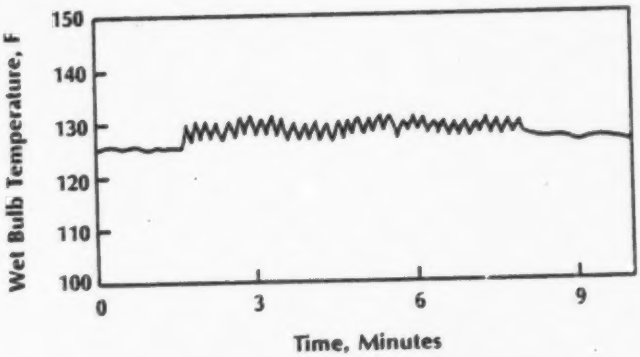


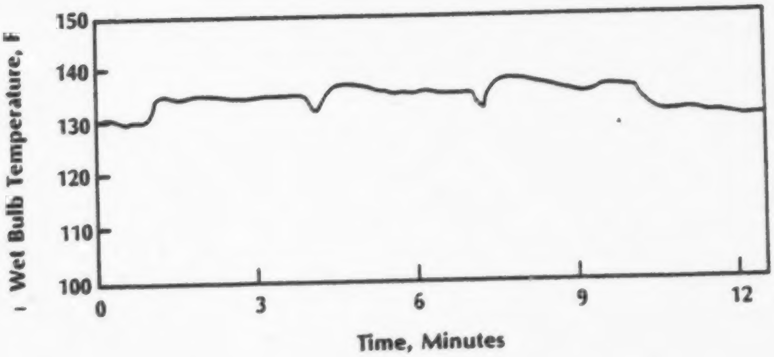
FIGURE 17 RIVERSIDE SPRAY DRYER RESPONSE TO RAPID LOAD FLUCTUATIONS



A. Wall Blowers
(Steam Flow = 6,000 to 8,000 lb/hr)



B. Economizer Blowers
(Steam Flow = 7,000 to 9,000 lb/hr)



C. Retractable Blowers
(Steam Flow = 10,000 to 12,000 lb/hr)

FIGURE 18 EFFECT OF STEAM SOOTBLOWING ON SPRAY DRYER INLET WET BULB TEMPERATURE

the steam soot blowers. The economizers and furnace wall blowers raised the wet bulb temperatures approximately 5 F. The retractable furnace soot blowers increased the wet bulb temperature by as much as 8 F. At low loads, soot blowing increased the wet bulb temperature by as much as 12 F at the spray dryer inlet.

Another phenomenon which occurs at Riverside is the stratification of dry bulb and wet bulb temperatures in the inlet duct to the spray dryer. This is caused by differing operation of the Unit 6 and 7 boilers. Figure 19 presents examples of stratified wet bulb temperatures at the spray dryer inlet during soot blowing with one unit off line. As shown on Figure 19, a 3 F stratification across the inlet duct and an 11 F increase over the initial wet bulb temperature occurs during peak soot blowing. With both units on line – one blowing the furnace walls and the other blowing the air heater – a wet bulb stratification of 5 F can occur across the duct. It was initially believed that the stratification would not be significant when the FGD system was treating flue gas from one boiler. Similar tests were performed at NSP's Sherburne County Unit 1 (750 MW). Measurements in the two air heater outlet ducts on this unit demonstrated that wet bulb stratification occurred across each duct and from one duct to the other. The stratification measured at Sherburne County Unit 1 corresponds with the number and location of the soot blowers in operation. The magnitude of the stratification at Sherburne Unit 1 was similar to that experienced at Riverside. It is important for successful low temperature operation of spray dryer systems that the fluctuations and stratifications in wet bulb temperature be quantified and monitored whenever possible.

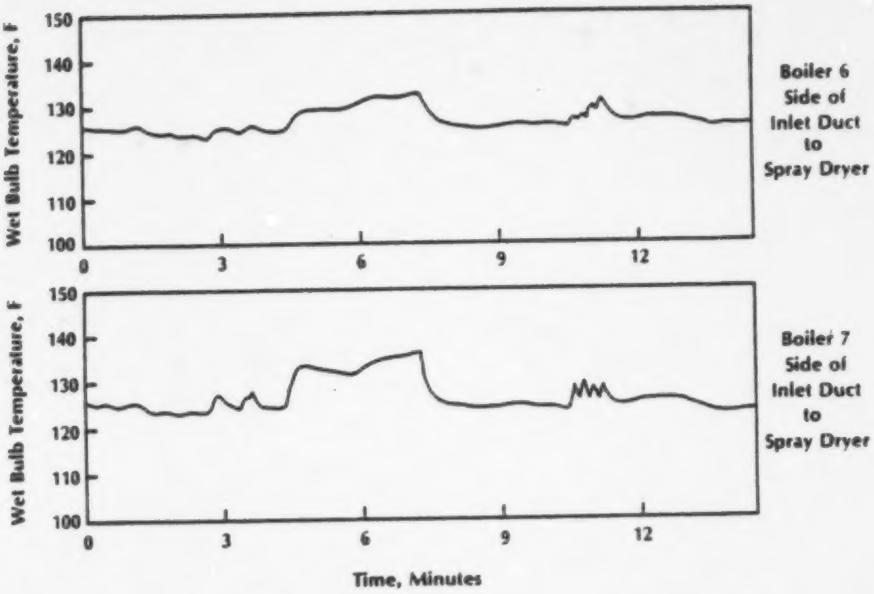
CONCLUSIONS

The Riverside spray dryer system has demonstrated the capability to provide compliance with New Source Performance Standards for SO_2 emissions. The average removal efficiency was 84.3 per cent for the 164 tests, including low lime and no lime additive tests. SO_2 removal efficiencies exceeded 95 per cent in 30 of the 164 tests. SO_2 removal ranging from 70 to 90 per cent removal was demonstrated for a variety of fuels.

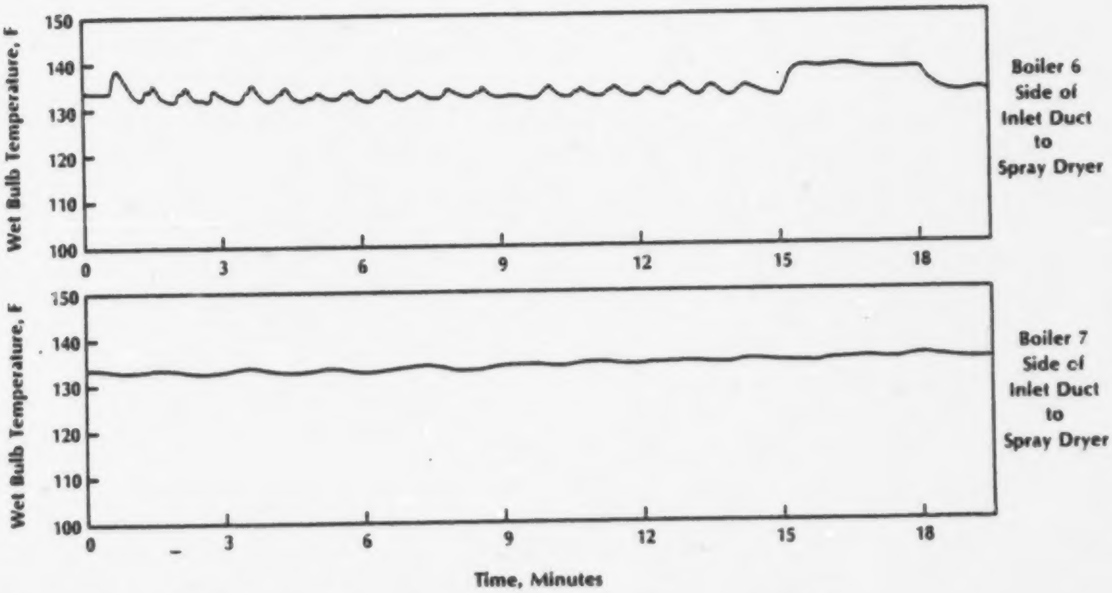
The spray dryer system is capable of drying the waste product to a powderlike consistency. The moisture content parameters in the data base indicate that the spray dryer system produced a relatively dry waste product in all of the tests. While operating the spray dryer during the two-year test period, upset conditions did occur occasionally which raised the moisture content of the waste solids to relatively high levels. The solids moisture content recorded in the data base represents typical steady-state utility operating conditions.

From an analysis of the performance test data collected during 1981 and 1982 at Riverside, certain overall correlations were apparent. A strong relationship between SO_2 removal, approach temperature, ash alkalinity, and lime stoichiometric ratio was observed. Other correlations between fabric filter SO_2 removal and spraydown temperature, and between spray absorber and fabric filter waste moisture content were apparent from the performance test data. Many other correlations which have been demonstrated on pilot size units were not observed from an evaluation of the overall data base.

The Riverside dry scrubber system responds well to changes in operating conditions. It was possible to maintain absorber approach temperature at an almost constant condition during changes in boiler operations. SO_2 removal efficiency typically changed with changing operating conditions; however, it was usually greater than required, such that it was possible to maintain compliance



A. Boiler 6 Off Line; Boiler 7 at Low Load



B. Boiler 6 Wall and Retractable Sootblowing; Boiler 7 Air Heater Sootblowing

FIGURE 19 STRATIFICATION OF FLUE GAS WET BLUB TEMPERATURE AT SPRAY DRYER INLET

almost continually. Variations in the absorber inlet wet bulb temperatures were observed during soot blowing operations. The maximum recorded wet bulb temperature variation during soot blowing was 8 F at full load and 12 F at low load. Sufficient margin must be maintained in the approach temperature to account for such operational variations.

The Riverside FGD system exhibits normal variations in performance consistent with a full scale operating system and, based on a broad range of data, is not as sensitive to variations in some parameters as are laboratory or pilot systems.

ACKNOWLEDGEMENTS

The authors gratefully acknowledge the contributions of Niro Atomizer Incorporated and Joy Manufacturing Company. The performance data presented in this paper were collected as a part of their testing efforts and were released to Northern States Power Company and Black & Veatch for independent evaluation. In addition, Northern States Power Company's Riverside operating staff is to be commended for their ability to endure the extensive testing program conducted at their installation.

DESIGN AND INITIAL OPERATION OF THE SPRAY DRY FGD SYSTEM AT THE MARQUETTE MICHIGAN BOARD OF LIGHT AND POWER SHIRAS #3 PLANT

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ABSTRACT

This paper discusses the design issues, design decisions, start-up, and early operation of the Spray Dry Flue Gas Desulfurization (SDFGD) system which went into operation at the Marquette Michigan Board of Light and Power Shiras #3 in May 1983. This forty-four (44) megawatt unit consisting of a rotary atomizer reactor, reverse air fabric filter, lime preparation, and reagent recycle system was engineered in the 1980-82 time period utilizing pilot plant and prototype industrial system results as a design basis.

The initial operation of the unit is discussed, as is the success of the scaleup from pilot plant to Commercial size boiler.

I. BACKGROUND

The development of the process and equipment design technology on which the Shiras #3 Spray Dry Flue Gas Desulfurization (SDFGD) System is based began in 1978 under a joint venture agreement with Anhydro A/S of Denmark. It merged the FGD process/equipment and particulate collection system knowledge of General Electric Environmental Services, Inc. with the spray atomization/drying knowledge of Anhydro into a single integrated system which was bench tested in Copenhagen, pilot tested at the City of Colorado Springs Martin Drake Station (CCS), and characterized in two years of testing (2, 4, 5) partially funded by the USEPA.

In the late 70's, the Board of Light and Power of the City of Marquette Michigan commissioned Lutz, Daily and Brain to design the totally new 44 MW Shiras #3 unit. The plant could have complied with lb/MBTU emission requirements for SO_2 via the compliance coal approach but was forced by the % removal clause of the 1977 Clean Air Act to include FGD in the plant design. For this size installation the emerging SDFGD technology provided the best balanced solution.

II. APPLICATION

Shiras Unit #3 is a 44 MW installation. Plant equipment includes a 415,000 pounds (188,244 kg) of steam per hour Combustion Engineering (CE) pulverized coal fired boiler fed by four CE coal mills using #2 fuel oil in the ignitors. The boiler utilizes steam soot blowing and is operated as a balanced draft system with single FD and ID fans and a CE regenerative air heater. The ID fan system, consisting of a Westinghouse Sturdivant radial tip centrifugal fan, American Standard fluid coupling, and 1250 HP (932 KW) /900 RPM (94 rad/sec) General Electric motor, was provided as part of the air pollution control system contract.

Design conditions utilized for the Air Pollution Control System definition are summarized in Table 1.

TABLE 1 DESIGN CONDITIONS

Air Heater Inlet Temperature	852°F (728°K)
Air Heater Outlet:	
Temperature	265°F (402°K) continuous 550°F (506°K) maximum 205°F (370°K) @ 25% MCR
Air Flow	226,000 ACFM (107 ACFM/sec) 723,000 #/Hr (91 kg/sec) 1500 PPMV SO_2 12% Vol. H_2O
Pressure	-4" W.G. (-1 kPa G)
Flyash	9075 #/Hr (1.14 kg/sec)

The basic SDFGD process is defined in the process flow diagram (Figure 1). In most cases, the flue gas goes directly form the air heater outlet to the spray atomizer/reactor. However there are retorfit applications where some flyash removal occurs as a result of existing cyclones or precipitators.

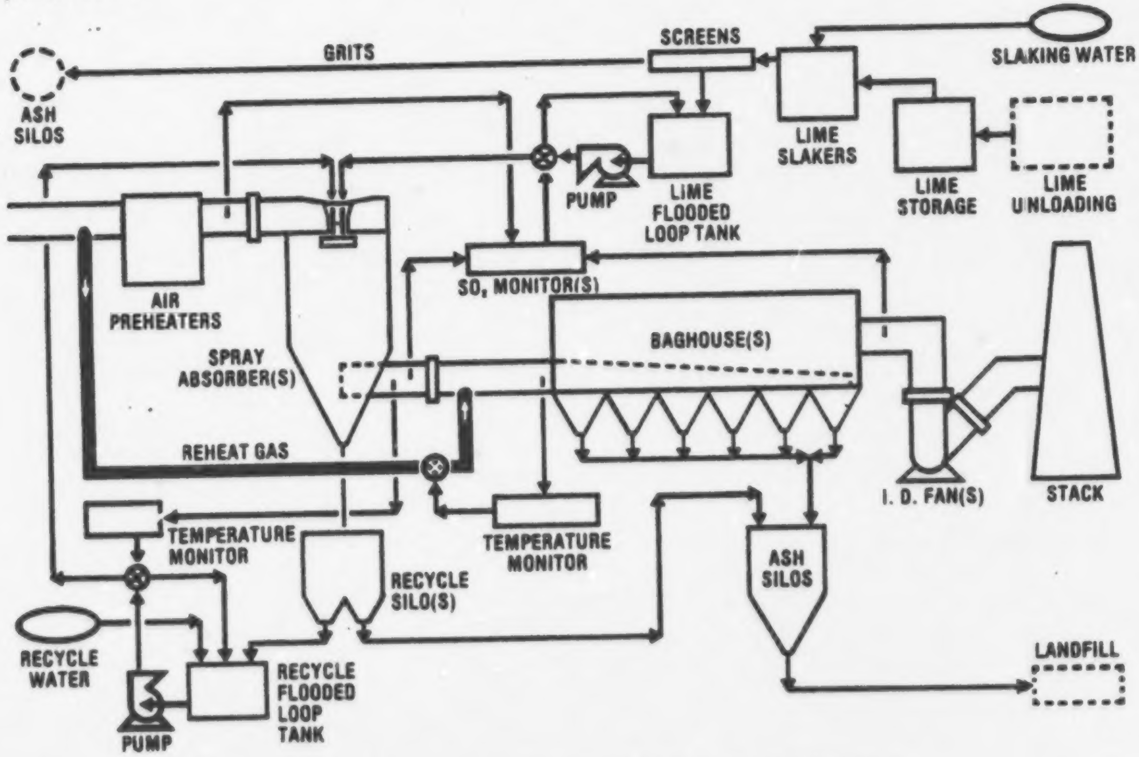

FIGURE 1. DRY FLUE GAS DESULFURIZATION SYSTEM

Aerodynamic devices in the chamber (inlet scroll, pre-swirl vanes, inlet annulus) impart appropriate axial velocities and angular momentum to the gas stream to insure uniform mixing with the reagent stream and cyclonic action that retains 40-55% of the dried product and flyash in the reactor vessel hopper. This total vessel cyclonic effect is unique to single atomizer/vessel systems, although other reactors (spray nozzle, multiple rotary atomizer/chamber) exhibit some collection of particulate in the vessel due to its "drop-out box" characteristics. The reagent slurry streams ("milk of lime" and recycle material) are fed to the centrifugal atomizer wheel with rates controlled respectively by system outlet SO_2 level and reactor vessel outlet temperature. The separate streams partially mix on the wheel and chemically react, but not to physical or chemical equilibrium because of the short (millisecond) residence times on the wheel. Equilibrium of these mixtures takes on the order of 5 to 20 minutes in controlled bench scale experiments. The design intent is to minimize the reaction between the two reagent streams to the greatest degree possible. The reagent stream passes from the central feed point on the wheel (Figure 2) to the tip as a thin Coriolis-force-induced liquid film in radial holes distributed about the circumference of the wheel. When the film reaches the tip, it is sheared off by the gas to form a droplet stream. The initial velocity of these droplets has a radial/tangential velocity ratio that depends on mass flow rate per hole and liquid viscosity but is always less than one. The droplet size distribution is affected weakly by film thickness and strongly by gas shear force (7, 8); hence atomizer wheel tip speed has the strongest influence on droplet size.

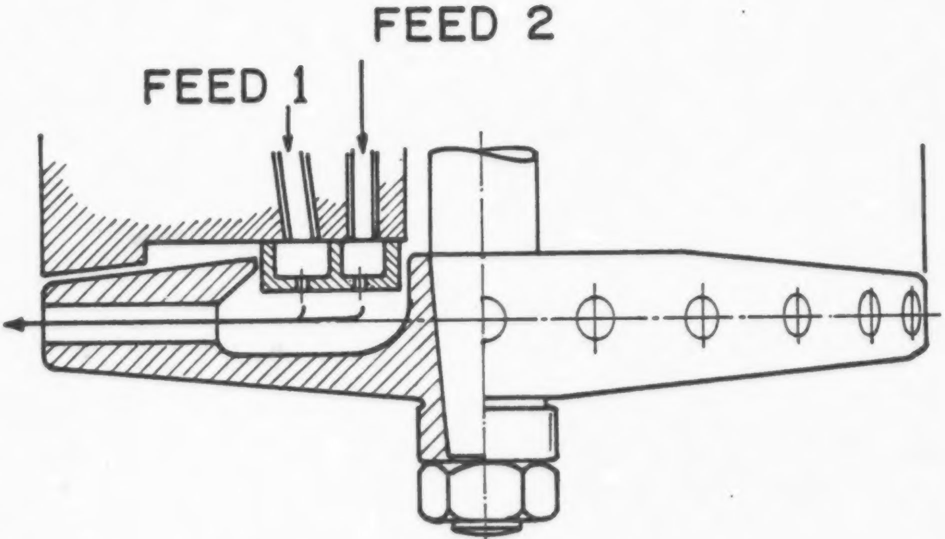

FIGURE 2. DUAL FEED ATOMIZER

The droplet streams form what looks like an umbrella to the stationary observer. The interaction of the inlet gas with this umbrella of droplets is a critical factor in controlling the uniform mixing of reagent and SO_2 laden gas. The drying of these droplets of highly alkaline (pH 8-12) liquid with suspended particles of (.5 - 4 micrometer) calcium hydroxide and (.5 - 30 micrometer) flyash into (10 - 50 micrometer) porous particle clusters proceeds through a complex series of surface chemistry and diffusion mechanisms (6, 9) to produce a relatively dry (1 - 3% moisture) particulate stream at the reactor outlet. The size of the resultant porous clusters depends on both the initial droplet size and the % solids in the feed slurry. For very high inlet temperature situations without recycle where the % solids in the slurry varies, this porous cluster size could be a significant variable.

The spray absorber outlet stream mixes with reheat gas of a quantity dictated by the desired particulate collector inlet temperature and proceeds to the particulate collector. Some further reaction occurs between the remaining gas stream SO_2 and the particulate material. In the case of a fabric filter, this additional reaction can capture 20-50% of the remaining SO_2; in a precipitator, the additional capture is in the 5-15% range.

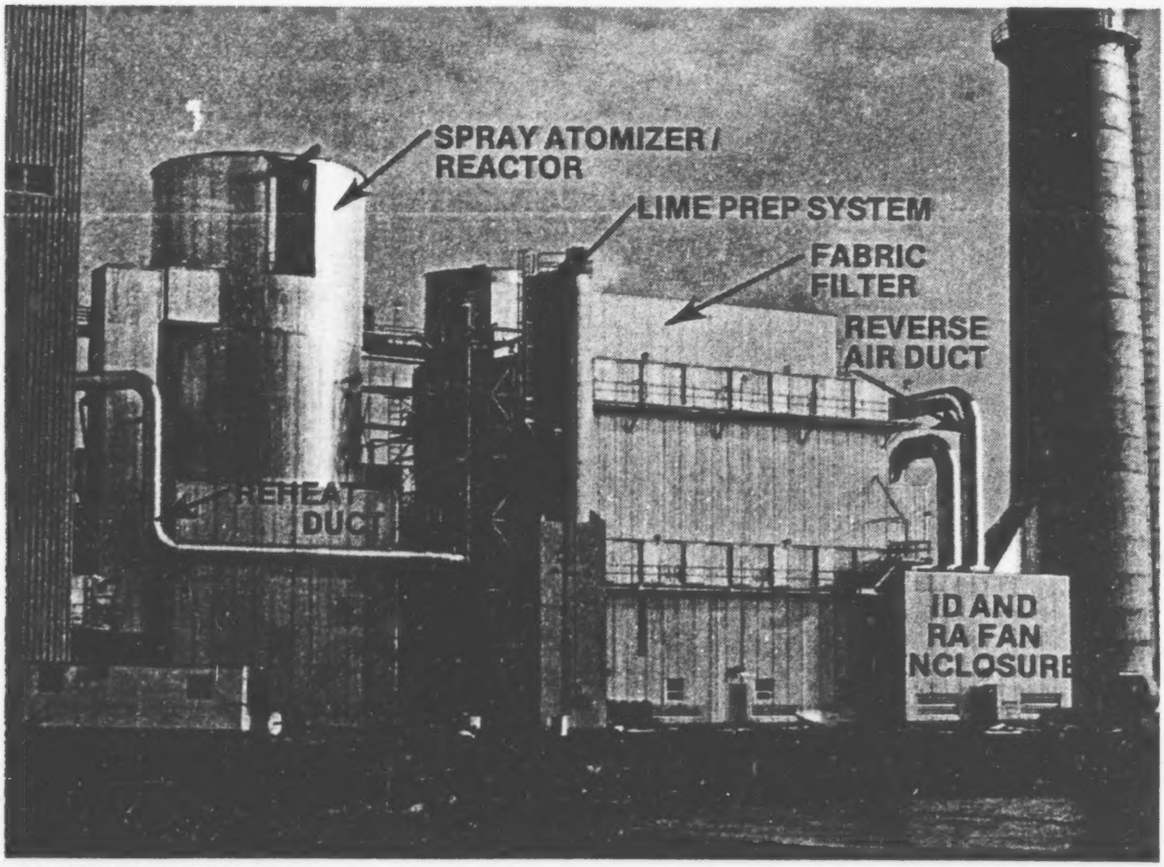

FIGURE 3. SHIRAS #3

The particle clusters normally have a central core of unreacted calcium hydorxide and a substantial amount of calcium carbonate produced by absorption/reaction of gas stream CO_2. While SO_2 is more reactive than CO_2, the high CO_2/SO_2 ratio in the gas stream counter-balances the reactivity difference and results in substantial calcium carbonate formation (4).

In addition, certain flyashes contain substantial quantities of alkaline material. Reslurrying the collected solids will get the residual alkalinity of the calcium salt/flyash clusters back into solutions while also providing relatively large nucleation sites for the fresh calcium hydorxide particles.

The physical equipment which turns the process flow diagram into reality is shown in the site photograph of the Shiras #3 unit (figure 3).

III. SDFGD SYSTEM DESIGN ISSUES AND INITIAL OPERATING EXPERIENCE AT SHIRAS NO. 3

There are many sub-system design issues which have to be resolved in the design of the SDFGD system. In this section, we will briefly define the issues and discuss how the designs have functioned during initial operation.

A. REAGENT PREPARATION

References 1, 4, 11 and 16 deal in detail with the effect of various reagent choices on the performance of the systems. Materials such as lime, limestone, sodium carbonate, Nahcolite, Trona, ammonia compounds, etc. were considered. For a variety of reasons, e. g., cost, performance, waste stability, and availability, soft-burned pebble lime has become the favored material for SDFGD systems. In the case of Shiras #3 the lime supply is high calcium with specified available calcium oxide content of 88%.

Lime is a generic term for a range of materials which vary in their C_aO content, may contain significant levels of M_gO, have varying amounts of "grits"*, and may have "hard burned" constituents (caused by surface melting in the kiln). This variation in composition causes a significant diversity of opinion in how the lime should be slaked to "milk to lime", a water slurry of $Ca(OH)_2$.

The slaking process is described chemically as follows:

$$CaO(s)+H_2O(1) \quad C_a\ (OH)_2\ (1)+15{,}300\ \frac{K - Cal}{Kg - Mole}$$

---

*"Grits" are composed of uncalcined limestone, kiln brick fragments, silica, alumina, ferric oxide, etc.

When this reaction is carried out with water of almost potable qualtiy and high calcium, soft-burned, pebble lime at a water/lime ratio of 3 to 4, its exothermic nature causes an almost explosive disintegration of the pebbles and a 70°-80°F (40-45°K) temperature rise in the water in four minutes resulting in a slurry of extremely fine (.5-4 micrometer) particles suspended in water. This slurry quality level is highly desirable for the SDFGD process.

The well known paste or detention slakers, which are simply mixing tanks with agitation, depend on this exothermic reaction to achieve the desired product. The larger "grits" are allowed to settle and are physically removed from the system via a combination of screw classifier, vibrating screen, and dilution water wash lowering the specific gravity of the lime slurry from > 1.3 to > 1.2. These commonly used slakers are simple, reliable, and relatively inexpensive.

When poor quality slaking water, hard burned contents, or MgO contents (almost always hard burned) are a significant factor, different slaker types such as attrition or ball mill are usually considered. Poor quality slaking water has dissolved solids which react with the surface of the lime pebbles, "blinding" their pores and preventing water infiltration, thus stopping the slaking process. Hard burning causes the same problem via a surface glaze effect, i.e., the pores left by the escaping CO_2 in the calcining process collapse due to overheat resulting in "blinding" pebbles. These more aggressive slaking methods use mechanical means to break up the surface and allow water contact with the internals of the pebbles to complete the slaking. Even with these techniques the quality of the reagent is not as good as that produced under the more ideal conditions. Even though these slaking systems can grind up the grits, they still should be rejected from the system. Since they are abrasive and require water for the transport of inert material, their retention results in an ineffective use of a key resource in this water limited process.

Shiras #3 uses a pre-assembled Clow-Coffman lime preparation plant. It consists of a pebble lime storage silo with an integral vent filter and dual vibrating bin discharges, redundant lime feeders, redundant lime slakers with agitators and grit removal screws, dual vibrating screen grit removers with associated screw conveyors, a lime slurry tank with redundant agitators, redundant slurry pumps, and associated flushing, drain and control systems. A major factor in the selection of this design was the fact that a similar system (without the operational redundancy) designed and built by the same vendor performed very well in the CCS pilot program (2). The lime slurry design range is 15-25% solids controlled by agitator current and mass/thermal balances.

In actual around-the-clock automatic operation we found that several small practical improvements could be made to the slaking system, and that one major issue had to be dealt with. Minor adjustments were made, such as the installation of a lime feed cutoff device that is activated if slaker water pressure drops below a given level. In a small power plant water supply line pressures have occasional wide fluctuations. It is better to stop slaking for a few minutes when this occurs, rather than depart from the optimal lime slaking temperature ranges (180°F to 190°F).

A more serious problem is that we found that one lime truckload out of a dozen had grit contents over 20%, most of which was unburnt limestone. This problem was later traced by the lime supplier to learning curve quality control problems caused by the change from a gas-fired to a coal-fired kiln. A large percentage of the limestone grit would leave the slaking system in the form of a fine suspension, and cause pluggage problems in the lime distribution system. It was found that a simple hydroclone was adequate to filter out the calcium carbonate, provided that one took into account the highly viscous, non-Newtonian nature of the lime slurry and over-sized the hydroclone accordingly.

B. SPRAY ABSORBER

The chamber, where the reagent and recycle slurries are atomized into fine droplets, where the reagent droplets mix with the incoming flue gas, where the absorption/chemical reaction of SO_2 takes place, and where a significant amount of the dried particulate is collected, is the most critical piece of process equipment. A cutaway schematic of a typical spray absorber is shown in Figure 4. Operational experience with hundreds of industrial spray drying applications provide the user with a range of cost-effective reliable alternatives. The major concerns that utility SO_2 service causes for this proven equipment is the abrasive and corrosive constituents in the inlet flue gas; abrasive chemically reactive slurries are not new to the spray dryer industry (13).

Reference 14 gives a credible review of the choice between two fluid spray nozzle systems and spinning disk rotary atomizer systems. The main factors in that choice are cost, power consumption, droplet size and variability control, reliability, and gas/droplet mixing. Those vendors who supply high horsepower rotary atomizers use one centrally located in each reactor of a multiple reactor system, treating up to 150 megawatts equivalent gas flow per reactor. This symmetry allows use of the vessel as a moderately efficient cyclonic collector. This can be particularly useful in protecting the downstream particulate collector from wet material in case of upset conditions. Our emphasis in this paper is on the single rotary atomizer system supplied by Anhydro A/S to GE for the Shiras #3 installation. The spray absorber/reactor consists of a low pressure drop gas disperser, belt-driven atomizer, and absorber chamber. The spray absorber must be carefully optimized so that the reagent slurry is atomized, contacted with the flue gas, and dried in a manner that promotes maximum capture of SO_2, minimum reagnt consumption, and low energy use, while maintaining stable and reliable plant operation.

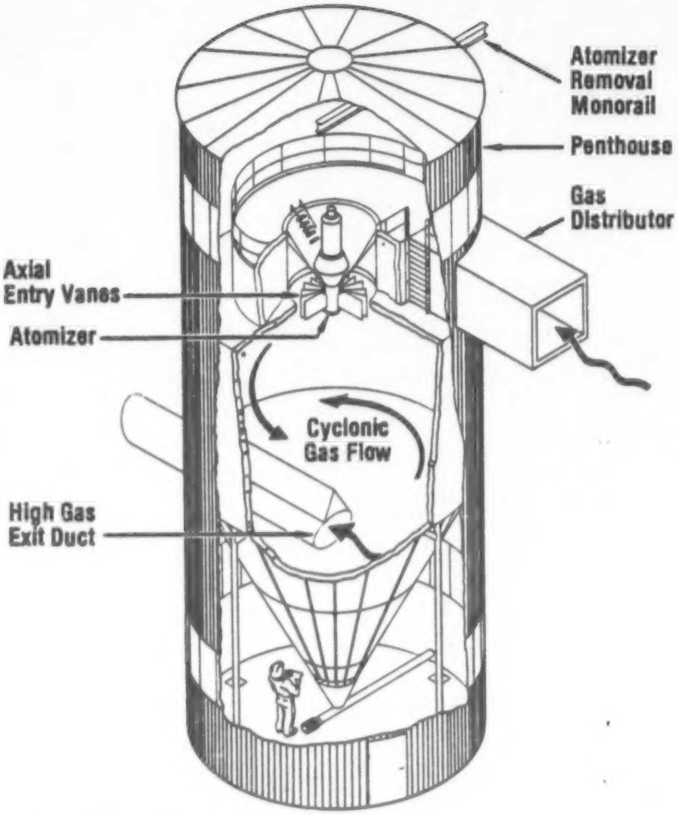

FIGURE 4. SPRAY ABSORBER

For the dry SDFGD application, we use a standard 400 mm tip diameter centrifugal atomizer designed to generate a uniform spray of fine (10-80 micrometer) droplets over a wide range of feed rates. For larger gas flows, the number of radial holes in the atomizer wheel is increased. The droplet size model used in our system design has been confirmed in detailed experiments at GE's Corporate Research and Development Center. Typical results over a range of tip speeds are shown in figure 5.

For the Shiras #3 atomizer, this means that the mass mean particle size is about 30 microns.

The two slurry feeds are piped independently (see figure 2) to a co-annular liquid distributor. In this way the two chemically reactive calcium hydroxide and recycle slurries are kept separate until fed on to the atomizer wheel.

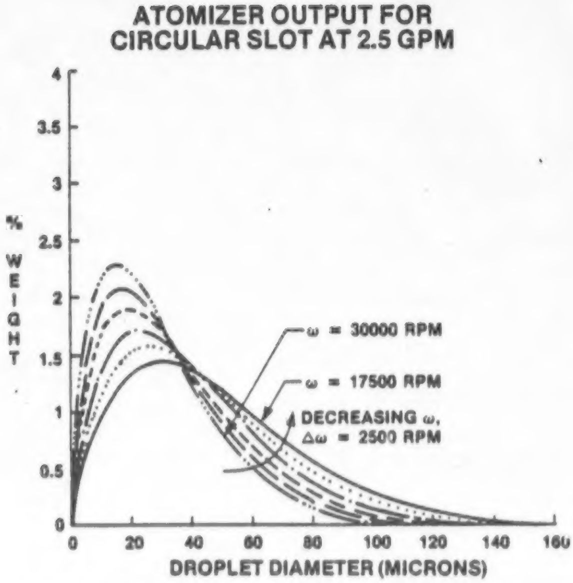
ATOMIZER OUTPUT FOR CIRCULAR SLOT AT 2.5 GPM

FIGURE 5. FOUR INCH WHEEL DATA

The slurries are introduced to a central cavity on the rapidly rotating atomizer wheel, and are induced by inertial forces to flow outward through radial passages in the wheel and then break off to form spherical droplets whose size range is chiefly governed by the viscosity and surface tension of the liquid and the atomizer wheel tip speed. As the droplets move away from the wheel and disperse into the gas stream exiting the gas dsiperser vanes, they form an umbrella-shaped spray pattern that is symmetric about the chamber axis and serves as the zone of initial contact between the reagent and flue gas. Figure 6 shows a stroboscopic photograph of one ligament of that umbrella.

The size of the droplets must be controlled to assure optimum reaction; small droplets are desirable because they provide a large surface area for mass transfer, but they must not be so small that they do not penetrate to the outer diameter of the inlet gas annulus.

A belt drive system, which makes it possible to change atomizer wheel speed by simply changing the belts and sheaves, is used on Shiras #3.

Because of the abrasive character of the slurries atomized in the spray absorption process, a wheel design is used which features silicon carbide inserts in the slurry passages. The inserts may be rotated as local wear spots appear to extend their useful life.

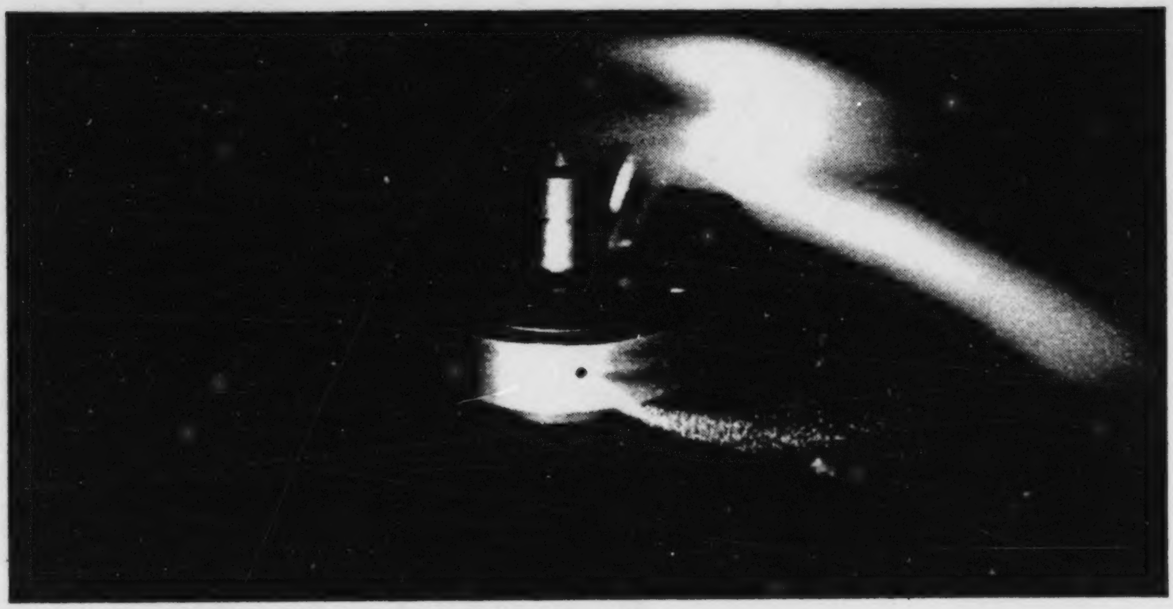

FIGURE 6. DROPLET STREAM STOP-ACTION

The Shiras vessel is 36 feet (11 meters) diameter, 71 feet (21.6 meters) tall and uses a 200 HP (150 KW) motor to drive the atomizer wheel at 7800 RPM (817 rad/sec) thru a custom designed belt/sheave system.

The absorber gas disperser is designed for high SO_2 removal without high pressure drop or flyash abrasion problems. Anhydro selected a top-entry vaned scroll-type gas disperser, which discharges an annular vortex flow of flue gas down into the chamber on all sides of the atomizer wheel. The gas disperser is equipped with vanes whose angle can be adjusted to obtain mixing of the spray and the flue gas. The initial gas rotation from the inlet scroll and vanes is in the same direction as wheel rotation which provides an increased cyclonic effect but slower deceleration of the droplet stream and larger droplets.

The absorber chamber must be sized in relation to the gas flow volume to assure that the slurry droplets will have adequate residence time (in most applictrions 8 to 12 seconds based on absorber outlet volume) in the chamber for the various stages of reaction with SO_2 and drying to occur. The chamber design also affects the degree of dropout in the absorber, which should be maximized to reduce the particulate loading to the fabric filter. The absorber is designed as an axial-entry cyclone to achieve approximately 50% dust dropout in the chamber.

The GE CR&D Center has done extensive coupled modeling of the gas flow, particle trajectory, drying, absorption and chemical reaction to provide a tool for correlating operating data and guiding future design optimization.

The model results indicate that the key phenomena occurring in the gas inlet/droplet injection region of the vessel is mixing of the gas and reagent; the bulk of the absorption process occurs after this initial gas/slurry mixing phase.

If the droplet/gas momentum ratio is too low, the gas at the outer diameter of the inlet annulus will see no reagent, i.e., bypass occurs. If the ratio is too high, a similar bypass core occurs at the inner diameter and the larger droplets get picked up by a recirculation vortex and get deposited on the wall of the atomizer. Since these ratios are constantly changing as boiler load changes, the Shiras #3 design divides the inlet into three dampered co-annular rings so that high gas momentum can be maintained as gas flow volume changes. The variable speed drive concept mentioned earlier can also be used to advantage in this turndown management system, i.e., smaller droplets at low gas velocities, bigger ones at high velocities.

A practical illustration of the importance of the droplet/gas momentum ratio was provided early in the Shiras #3 startup. Persistent boiler soot blowing system problems resulted in weeks of steady-state gas temperatures into the spray absorber of between 350°F and 410°F; far above the design value of 265°F. The GE/Anhydro atomizer had the reserve power and pumping capacity to increase slurry flow and maintain constant approach temperature. However, this resulted in a significantly higher droplet/gas momentum ratio, and a much broader reaction umbrella, which caused minor solids deposits on the vessel walls and roof to be built up over a one month period. As soon as this was seen, the setting on the gas distribution vanes (the "wing vanes") was altered to permit a more vertical gas flow into the vessel, thus reducing droplet/gas momentum ratio. After another month of operation, the vessel was inspected, and the roof and walls were free of deposits.

C. PARTICULATE COLLECTOR

In a SDFGD system the spray atomizer/reactor is primarily an SO_2 capture device but doubles as a particulate collector. Not to be outdone, the fabric filter (or precipitator in some cases) particulate collector also doubles as an SO_2 absorber. Effective operation in this secondary mode requires operation of the fabric filter, within 20-50°F (11-28°K) of the gas dewpoint. Although many utility baghouses operate that close to the sulfuric acid dewpoint, the quantities of liquid potentially produced are so different that special design attention is rquired in the SDFGD case, particularly in the areas of reheat control, internal/external insulation, and leakage control.

References 10 and 15 thoroughly discusses the pros and cons of precipitators and fabric filters and the special design considerations for each.

In summary, the superior FGD absorption contribution of the fabric filter makes it the logical choice in high efficiency applications or in any situation where lime costs are high and operating cost savings are seriously considered.

The Shiras #2 unit is a conventional eight (8) compartment custom GE reverse air design. It uses 168, 35 foot (10.7 m) long, 12 inch (30.5 cm) diameter, acid-resistant-coated, fiberglass bags per compartment. All fabric filter dampers are of the low leakage poppet type except the bypass which is a pressurized double louver selected by the LD&B engineers because of their excellent experience at the City of Colorado Springs Martin Drake Plant. The fabric filter system is a conventional design with redundant reverse air fans, programmable controller control of the cleaning cycle in a variety of operator selected modes (manual, time, P batch, P distributed, etc.), and a centrally manifolded inlet, outlet, R/A duct system.

With the atomizer in operation, the fabric filter design conditions are based on gas at an inlet temperature of 175°F (325°K) and a gross air to cloth (A/C) ratio of 1.49. With the atomizer off line the inlet temperature is 265°F (402°K) and the gross A/C is 1.62. Note that the design dust loading changes substantially between these two situations going from 6.0 to 4.7 grains/ACF. In actual operation this inlet gas tempreature to the system has been as high as 360°F (455°K).

Baghouse operation at Marquette has been exceptionally trouble free.

Operation has been at approach temperatures between 25 and 60°F. There have not been any occasions of condensation on the baghouse walls or the bag filter cakes (all baghouse compartments are kept on line regardless of load level). There has not been any build up in average baghouse pressure drop in over four months of operation.

Most critical and interesting of all, the frequency of the baghouse batch cleaning cycle actually drops when the spray absorber is brought online. Typically, if the baghouse has been cleaning 50% of the time while the atomizer is being inspected, it will drop to cleaning only 33% of the time once the atomizer is restarted.

Temperature drop across the baghouse is minimal (under 5°F) when the spray absorber is operating.

D. REHEAT

The evidence is conclusive that operation with reactor discharge temperatures as close to adiabatic saturation as possible optimizes the lime use - collection efficiency relation. However, temperature variation within the reactor due to non-uniform mixing of reagent and flue gas, wall deposits, reactor hopper pluggage and ductwork condensation, etc., constrain this optimization. Recognizing that reactor and particulate collector constraints on approach temperature are different, most designers have made provisions for reheating the gas downstream of the reactor. Several methods have been considered for accomplishing this reheat, for example:

a. Bypass of air heater inlet gas.

b. Bypass of air heater outlet gas.

c. Recirculation of fabric filter outlet gas through a water/steam - gas heat exchanger.

d. Injection of ambient air through a water/steam-air heater exchanger.

The first two cause similar problems. Because they contain particulate and SO_2, clean-up of the reheat system is required. If the reheat injection is upstream of the fabric filter, this clean-up is automatic. However, since the fabric filter is a low efficiency SO_2 device (the precipitator is even lower) which degrades in efficiency as its inlet temperature is increased, these types of bypass have a "double kicker" effect on required reactor efficiency.

Of the two, airheater inlet gas is theoretically more effective since smaller quantities are required for the same amount of reheat. However, the system heat rate effect is worse. The pilot work at Colorado Springs (12) raised some serious doubts about the validity of this theoretical conclusion which warrants further evaluation. The issue may be related to heterogenous mixing but even that is not clear.

Methods c and d use a scrubbed clean gas for reheat. They not only eliminate the "double kicker" (only the temperature degradation of baghouse SO_2 efficiency remains) but also provide the option for reheat injection at the baghouse outlet (eliminating the temperature degradation effect). This option assumes that reheat was not for baghouse protection purposes but for ductwork, stack, and plume buoyancy purposes. The negative factors in these approaches involve recirculation fan horsepower and water/steam BTU requirements.

In spite of the concerns raised about the theoretical conclusions by pilot results, Shiras #3 uses the air heater inlet bypass approach which requires a reactor efficiency of 80% to provide a system efficiency of 83% with 20°F (11°K) reheat. Air heater outlet reheat would require the reactor efficiency to be higher than the system efficiency as indicated in figure 7. For example a system efficiency of 80% would require a reactor efficiency of 100% if 22°F (12°K) reheat was required. This dramatically shows the impact of the selected reheat approach, particularly with the system design point on the high stoichiometry/ efficiency portion of the performance curve where lime savings more than counter balances heat rate increases.

Operation of the reheat system at Shiras #3 has been routine.

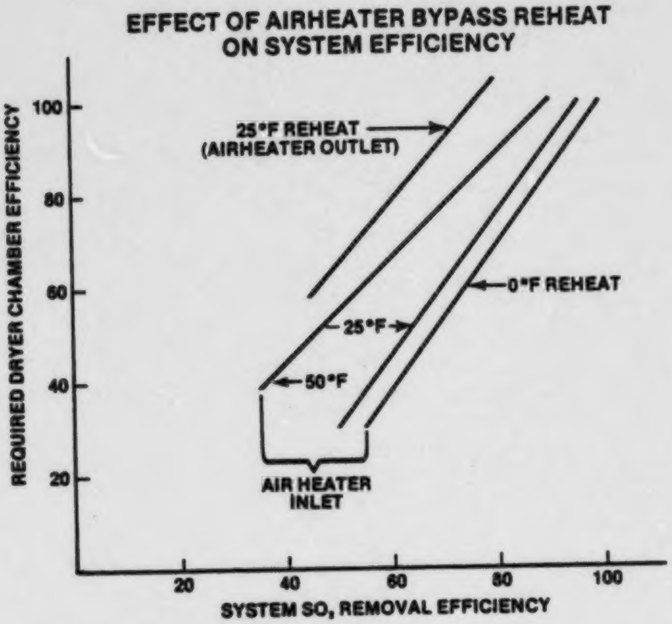
EFFECT OF AIRHEATER BYPASS REHEAT ON SYSTEM EFFICIENCY

FIGURE 7. REHEAT EFFECTS

Baghouse approach temperatures have been varied between 25°F and 60°F, and reheat has been used for the upper end of the temperature range.

E. CONTROLS

The important issues in DFGD system controls are:

a. Reagent feed control

b. Temperature control

c. Turn down control

d. System response to transients

The other control aspects such as lime slaker control, recycle slurry generator control, and baghouse control are not unique to SDFGD and are not covered in this paper.

Each design deals with the specifics of these control issues in its own way. Therefore, our focus will be generic with the specifics relating to Shiras #3.

It is clear that the reagent feed control strategy should be to maximize solids concentration to the limits permitted by solids drop-out and abrasion consideration and to maximize the ratio of recycle/fresh "milk of lime". In our system design we use the GE patented (3) Two Loop Control system which automatically optimizes system efficiency/stoichiometry. The lime slurry is prepared and stored at a solids concentration substantially higher than would normally be required to achieve the required absorber SO_2 removal level. For an optimum outlet temperature to be reached, the lime slurry is diluted with recycle slurry on the atomizer wheel just prior to atomization.

The first of the two automatic loops regulates the flow of lime slurry to the absorber, based on the fabric filter outlet SO_2 level, by modulating the position of the lime slurry control valve. SO_2 levels are also monitored at the absorber inlet and outlet using a Dupont extractive system.

The second loop controls recycle slurry flow based on approach to adiabatic saturation temperature at the reactor outlet.

Changes in the adiabatic saturation temperature due to steam soot blowing, coal moisture changes, ambient moisture changes, etc., are provided to the system via feed forward controls and/or operator input. Automatic determination of adiabatic saturation temperature is conceptually attractive but has been inhibited by the unavailability of proven reliable sensors. Although not part of the system design contract, we will evaluate three candidate sensors for this service at Shiras #3. The sensors are all extractive and subject to particle contamination if used at spray dryer inlet or outlet. It is conceptually possible to use these devices at the baghouse outlet but the increased time delay substantially reduces its value. Also, use at the outlet would be questionable because of the variable moisture absorption characteristics of the bag cake.

In operation at Shiras #3, it has been found the Two Loop Control system regulates the temperature leaving the spray absorber vessel to with + 3°F during normal operation, and to within ± 7°F during rapid boiler excursions.

For multiple (3 or 4) vessel systems, this range is enough to permit system turndowns of 40-50% and excess flows to 20% via vessel dampering and individual vessel shutdown. In the single vessel Shiras #3 design, the turndown is accomplished via variable inlet geometry in the vessel. The geometry variation involves dampering of flow in three co-annular inlet rings and variable swirl vane angle. Variable atomizer speed will also be tested, although not part of the basic Shiras design. The system allows variation of the gas/slurry momentum relations in the inlet area which dramatically affects system performance.

Turndown control is complicated for the single vessel Shiras design. Pilot data indicates that operation of a reactor vessel in the + 20% gas flow range is practical and has no more than a 5% - 7% effect on stoichiometry for a given efficiency.

System control dynamics issues are related primarily to the time delays in sensing the effect of slurry feed on SO_2 levels at the fabric filter outlet and temperatures at the reactor outlet and fabric filter inlet and the slow read rate of the SO_2 monitor. If not considered properly, system interaction with these loops could lead to excessive hunting, overshoot, and offset in the control system. To better understand the dynamics of the system, GE performed a computer simultion of the Marquette design. (17)

Operating experience has proven out that with 20 second effective response times on the recycle slurry loop and 15 minute response times on the lime slurry loop, that hunting does not occur.

F. MATERIAL HANDLING

The major issues in this area are the handling of potentially wet product from the absorber hopper and the sourcing of recycle material. Based on pilot experience (1,5), GE favors mechanical conveying (screw, drag chain, etc.) of reactor discharge material in combination with multiple assists for hopper clearage. These include vibrating bit bottom, poke holes, and an auxiliary side removal system from the hopper. In order to ensure sufficient recycle solids, reactor hopper and/or fabric filter hopper discharge should be available.

Because the Shiras #3 ash collections system was combined with upgrades on units #1 and #2, the City of Marquette purchased a pneumatic system which combined pressure and vacuum portions.

Operating experience has been that the pneumatic system needs a grinding system to break up occasional loose clumps, and such a device is being added to the ash removal system.

Another practical consideration is that although the solids hopper catch handles like a dry, western flyash, the boiler gas is brought to within 20 to 40°F of its dewpoint. Thus care must be taken to thoroughly heat trace and insulate pneumatic vacuum systems, since they will entrain some boiler gas which must not be allowd to cool and condense. Another solution is to use heated (250 to 300°F) ambient air in the pneumatic system. The former was done at Shiras #3 for the baghouse hoppers, and the latter for the spray absorber hopper.

G. MATERIALS OF CONSTRUCTION

Because of the non-corrosive nature of the gas stream down-stream of the reactor inlet, assuming condensation is avoided, carbon steel is used throughout the reactor, ductwork, and baghouse. Since the centrifugal atomizer outer shell is colder than the inlet gas and subject to untreated inlet gas, 300 series stainless is utilized in this protective cover.

Slurry piping is unlined heavy wall carbon steel; valves are rubber lined. Acid resistant coating was used on the fiber glass bags because its superior "fiber wetting" characteristic will protect the bag material from wet alkali attack; there is no acid attack concern.

So far the only sign of erosion in the entire system has been the expected wear at the atomizer wheel ceramic ports.

H. INSULATION/LEAKAGE/ T

Because of the low approach temperature/minimum reheat sytem design, there is a real potential for localized gas condensation on duct surfaces, reactor or fabric filter walls, etc. To minimize the risk, extra thick insulation (8 inch (20.3 cm) on reactor, 6 inch (15.2) plus air gap on fabric filter) was employed, special leak tests were run on the reactor and fabric filter, and purge dampers were not included on the fabric filter. One feature that might have been desirable was internal wall insulation in the fabric filter to avoid compartment off-line cold wall effects. There are concerns about condensation under the insulation in this approach, but it could have design merit if impervious insulation is used. Marquette doesn't have internal wall insulation. The fact that all compartments will be on line at all times (except for on-line maintenance) reduces the cold wall condensation concern. Units that have had this problem were running continuously with off-line compartments in an attempt to minimize fabric filter gas temperature drop.

Operating exprience has been that (with the exception of the vicinity of one leaky door gasket) wall condensation is not occuring, and that temperature loss across the system is under 5°F.

I. PERFORMANCE

Design performance requirements for the Shiras #3 system are:

1. 80% SO_2 removal at all boiler loads from 20 to 100%.
2. Outlet particulate of .005 grains per ACF or 99.5% weight efficiency which ever is less stringent.
3. Lime use not to exceed 2874 #/Hr (.36 kg/sec) at design conditions.
4. System pressure drop of 12.4" W.G. (3.1 kPa G)
5. Gas temperature at stack inlet of 165°F.
6. Bypass reheat of 3,600,000 BTU/Hr (1055 KW) based on 852°F (729°K) gas.

In September 1983 compliance test of all the above objectives were met, even though the system was not yet optimized, e.g., air preheater exit gas temperatures were from 70 to 80°F above the design condition.

A comparison of the compliance test data with 1980 data from the 2 MW demonstration facility at the City of Colorado Springs, Martin Drake No. 6 plant, shows that for the same stoichiometries, SO_2 removal efficiency was 10 to 20% higher in the 44 MW system than in the 2 MW system. Thus, scale-up has been properly applied in the Marquette system.

The overall conclusion from the first four months of spray absorber performance is that it is possible to operate at SO_2 removal efficiencies between 70 and 90% simply by varying lime stiochiometry. This performance, as well as the lack of scale-up problems, is attributed to the Marquette design which was supported by the R & D effort at General Electric's Schenectady laboratories.

J. ON GOING PROGRAMS AT MARQUETTE

For DFGD vessels handling from 100MW to 150MW of boiler gas each, 500 HP to 800 HP atomizer motors are required. Radial bearing loads for either belted or spur-pinion gear systems become excessive for these high horsepower drive trains, and bearings B-10 lives become unacceptibly short (under 10,000 hours). As a result, General Electric has developed a Direct Drive System which uses a voltage inverter to operate a vertical AC motor at power frequencies other than 60 Hertz, so that the motor rotates at the atomizer speed. A flexible metal coupling is used to transmit the torque from the rotor to the atomizer.

A Direct Drive system (shown in Figure 8) has been successfully operated for several months at the General Electric Turbine Technology Laboratory in Schenectady, New York, and is currently being installed at Shiras #3 for field testing.

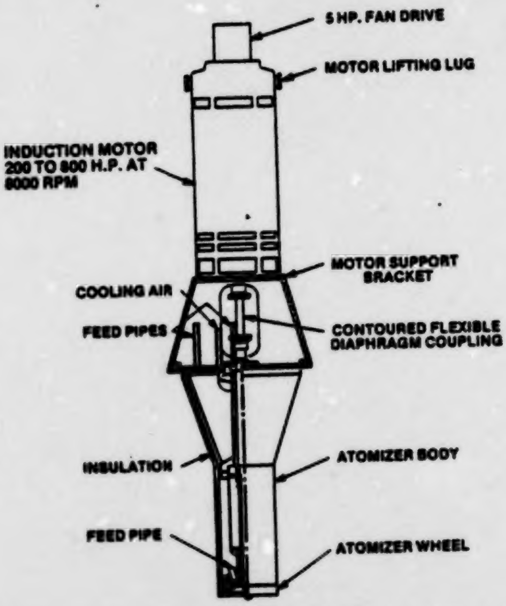

FIGURE 8. DIRECT DRIVE ATOMIZER ASSEMBLY

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START-UP AND INITIAL OPERATING EXPERIENCE OF THE ANTELOPE VALLEY UNIT 1 DRY SCRUBBER

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ABSTRACT

The first competitively bid and awarded utility dry scrubbing system utilizing lime as the scrubbing reagent was for Basin Electric's Antelope Valley Station Unit No. 1. Awarded in 1978 to Joy Manufacturing with Niro Atomizer as the major subcontractor, the system was scheduled to start-up in 1981, however, due to reduced load growth, was delayed until this year.

The dry scrubbing system treats flue gases from a 435 MW lignite fired boiler and consists of reagent preparation equipment, five spray dryer absorbers, and two fabric filters.

Initial operation on coal began in May, 1983 and commercial operation of the system is scheduled for July, 1984. This paper will review the start-up procedure, any problems which have developed thus far and how they have been handled, as well as the results of the operation of the system to date.

INTRODUCTION

In September, 1978, Basin Electric awarded a contract for a pollution control system to the Western Precipitation Divison of Joy Manufacturing with Niro Atomizer as a major subcontractor. The contract was for the supply and erection of an SO_2 dry scrubbing system and baghouse for Unit #1 at Basin Electric Power Cooperative's Antelope Valley Station near Beulah, North Dakota. Major equipment included in this contract were the reagent unloading and storage system, reagent preparation equipment, five spray dryers, mechanical conveyors for a recycle system, two baghouses for particulate collection, and a computer control system for the operation of the dry scrubber.

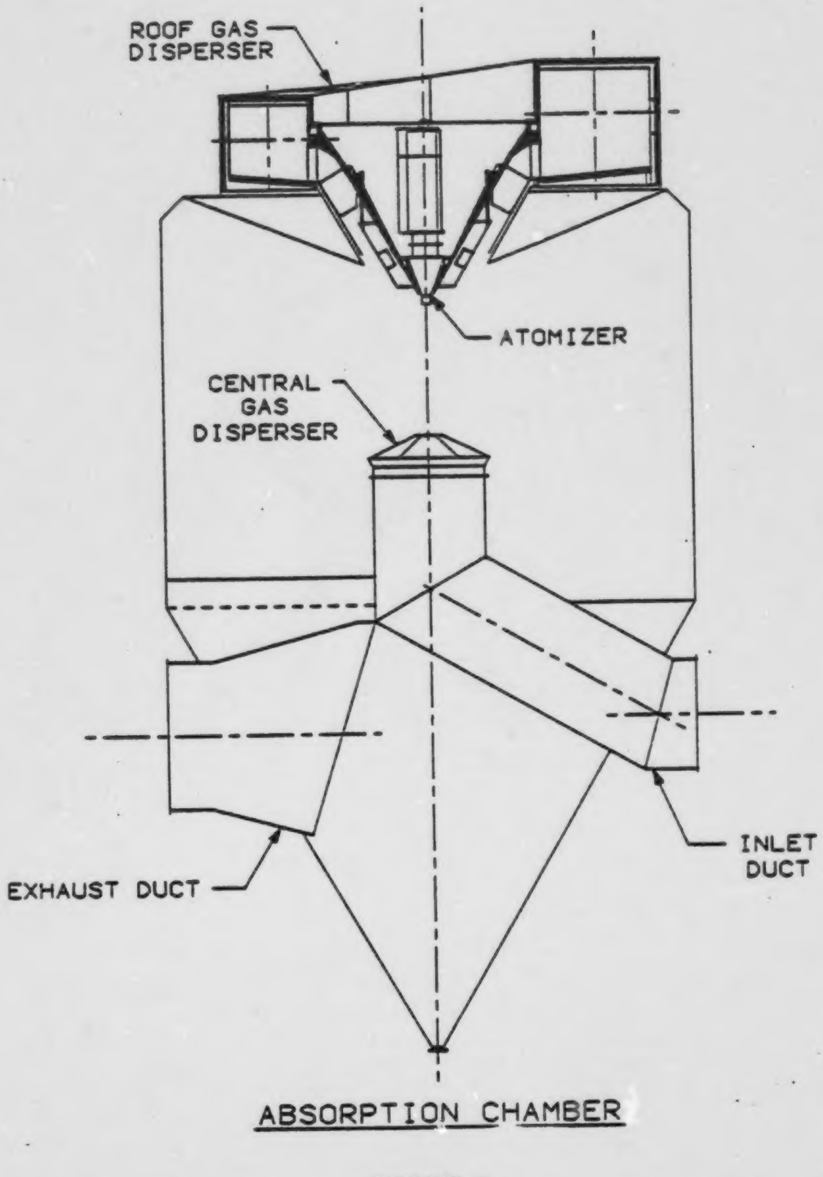
Engineering commenced in November, 1978, with field mobilization and the first equipment arriving on site in July, 1979. Field erection activities lasted from then until May, 1982. The plant was originally scheduled for commercial operation in 1981; however, due to lower than anticipated regional power demands, the actual plant start-up did not occur until May, 1983.

Joy and Niro start-up personnel arrived on site in November, 1982, and began pre-start check-out of the spray dryers and baghouses, including checking motor rotation, programming of the computer control system, etc. The actual system start-up began in May 1983, when the Combustion Engineering 435MW lignite boiler was first fired with coal, although the first flue gas to the system occurred on January 13, 1983, when the boiler was fired on oil. Start-up and testing lasted until the first week of October, 1983, when the performance testing to verify operating parameters as guaranteed (lime consumption, pressure drop, power consumption, etc.) was successfully completed.

SYSTEM DESCRIPTION

Dry Scrubber

The Antelope Valley Station dry scrubber consists of five 46-foot diameter spray dryer absorbers. An inlet manifold equally distributes the boiler flue gas to each of the five spray dryers (Figure 1). Gas is distributed between the roof and central gas dispersers of each spray dryer. Each dryer is equipped with a single, F-800 model rotary atomizer, which is powered by a flange mounted, vertical shaft 700 hp motor.



ABSORPTION CHAMBER

FIGURE 1

The full design gas flow of 2,055,000 ACFM (at 307°F) can be handled by any four of the five spray dryers with the number of modules selected for service based on boiler load. The inlet and outlet ducts of each absorber have guillotine isolation dampers to isolate individual modules when they are not in service. The scrubber computer control system signals the operator when spray dryers need to be brought on or off line (due to changes in gas flow), and the operator then initiates the appropriate sequence.

The slurry feed to the atomizers is prepared in a continuous feed preparation system with lime, water, and recycle materials controlled to match the system demands. Two complete lime slaking and feed preparation trains provide redundant equipment, maximizing system availability. Each slaking train is sized for 100% of the required lime consumption at the average performance conditions. The mixing trains are each sized for 100% capacity under maximum operating conditions.

Figure 2 shows the overall process scheme. Pebble quick lime is pneumatically transferred from the railcar/truck unloading facility to the 3,000 ton capacity silo and from there, conveyed pneumatically to a 100 ton capacity lime bin. Weigh belt feeders meter the lime from the lime bin into the two Denver Equipment ball mills, each sized to slake 6 tons/hr of pebble lime. The ratio of the treated water and pebble lime fed into the mills is controlled to maintain a 35% solids slurry within the mill. Slaked lime slurry leaving the ball mills is diluted to 20% solids, and classified in spiral classifiers to 100 mesh. The grits separated by the classifier are recycled back into the ball mill. The classified lime slurry is pumped by rubber lined pumps to the mixing tank, where baghouse and spray dryer recycle material along with dilution water are added. The feed slurry in the mix tank is continuously pumped to the feed tank, which overflows back to the mix tank.

From the feed tank, the feed slurry is pumped through individual supply lines to head tanks located above each spray dryer module. There is a continuous overflow from the head tanks back to the feed tank. The purpose of the head tank is to supply slurry to each atomizer feed control valve at a constant pressure. The control system regulates the amount of feed delivered through the control valve to the rotary atomizer to maintain a constant outlet temperature at each spray dryer exit.

Lear Siegler SM810 SO_2 monitors, located at the scrubber

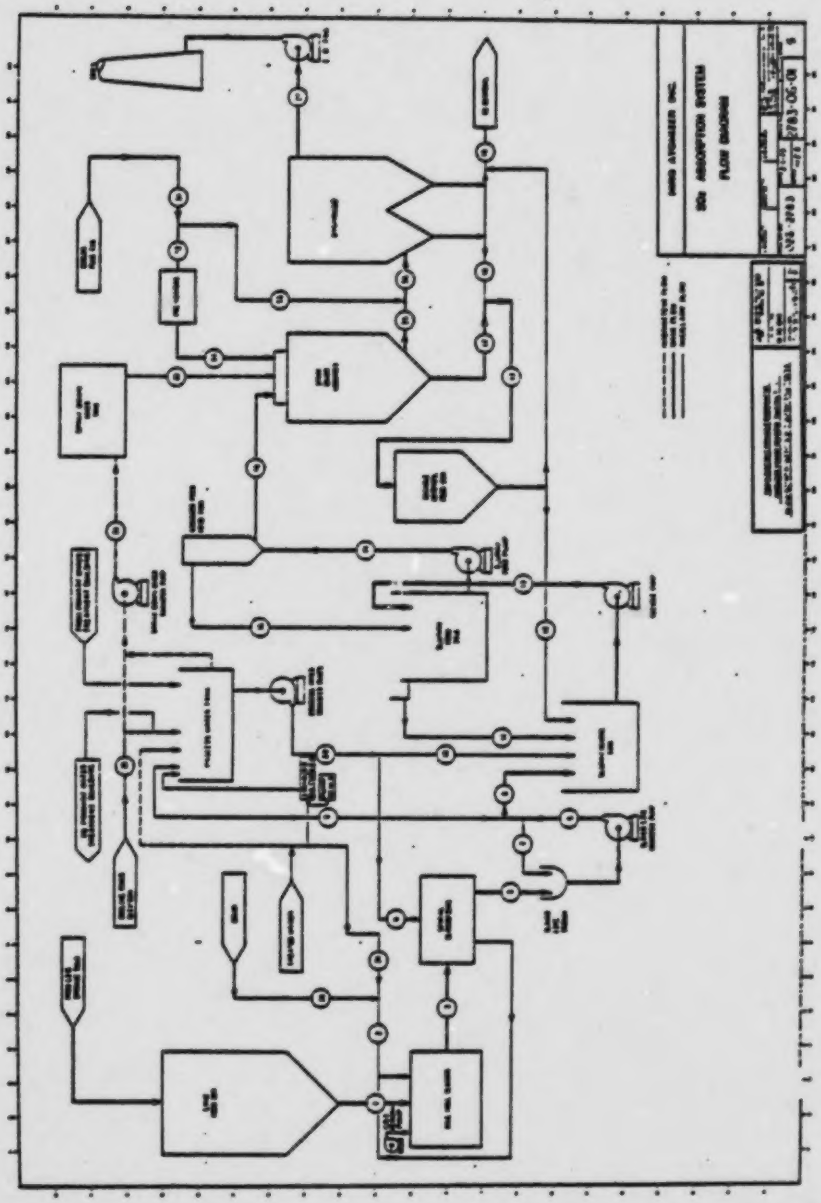


FIGURE 2 - FLOW DIAGRAM

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inlet duct and in the stack, feed signals into the scrubber controls (Honeywell TDC-2000 computer system). Boiler flue gas flow, temperature, and SO_2 content are continuously monitored and the feed preparation system inputs are varied continuously to maintain a 35% total solids content while obtaining the desired SO_2 removal at the required stack temperature.

At the outlet of each spray dryer is an inclined mechanical drag link conveyor to transport the dried powder and ash collected in each absorber. Gas exits from each spray dryer via a side discharge duct along with entrained flyash and spray dried material. A portion of the particulate in the gas stream drops to the bottom of the spray dryer and on to the drag conveyor for reuse in the feed system. The major benefit of this two point collection concept is that if wet material should exist in the absorber due to an upset condition, it will fall to the bottom of the chamber and be collected by the drag conveyor, while the flue gas will exit from the side discharge duct in the conical section of the chamber. Therefore, a clear flue gas path will always exist.

From the five spray dryer conveyors, the collected recycle material is deposited onto two horizontal collecting conveyors, and from the collecting conveyors, the material dumps into a single vertical lifting conveyor which discharges into the recycle bin. A portion of the ash collected in the baghouses is pneumatically conveyed to the recycle bin to insure that an adequate supply of recycle powder is available to support process requirements. The required amount of recycle powder is metered by a variable speed drag link conveyor into the mix tank along with lime slurry and dilution water to make up the final feed slurry. The use of recycled powder in the process is to reduce the amount of fresh lime needed for SO_2 scrubbing, and to enhance the drying aspects of the overall spray absorption process.

Baghouse

The two particulate collectors installed at Antelope Valley for collection of flyash and spray dried product are Western Precipitation reverse air baghouses. Each baghouse consists of two parallel rows of seven compartments for a total of twenty-eight compartments. The compartments are connected to the common inlet, outlet, and reverse air manifolds which are located on the centerline of each baghouse (Figure 3).

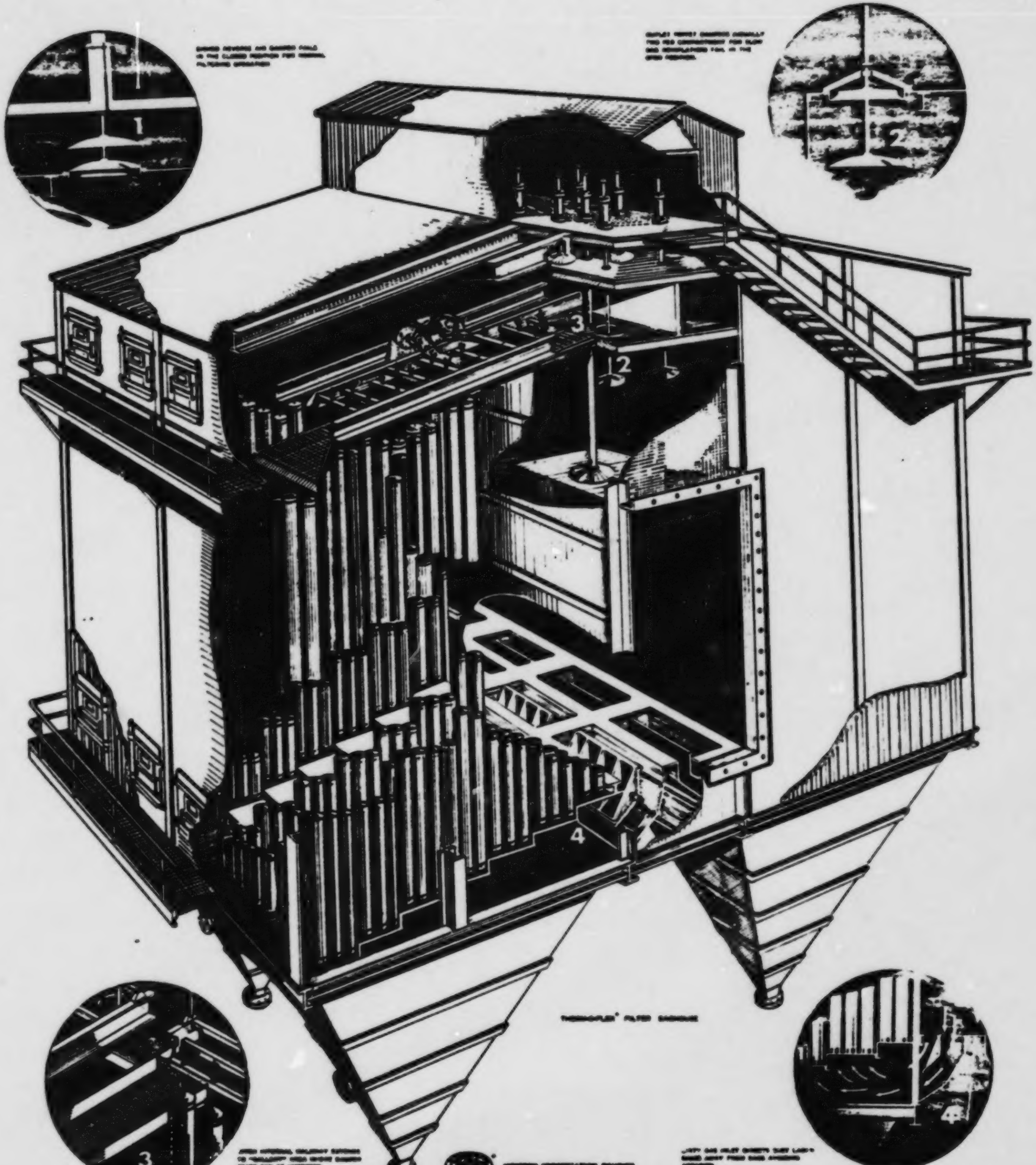

FIGURE #3 - BAGHOUSE DRAWING

Each compartment contains 288 filter bags. The bags have a 12-inch diameter and are 35 feet long. The bag type is Menardi-Southern 601T Teflon coated fiberglass. At the maximum design gas volume of 1,894,000 ACFM, the effective filter ratio is 2.19. The design operating temperature is 192°F. The bags are suspended under spring tension from a hanger bar at the top of the compartment and retained at the bottom by a flared thimble. The thimble floor is "stepped" down towards the baghouse outer walls. Beneath each compartment are two pyramidal hoppers, each equipped with an access door, poke holes, a hopper heating system, and a nuclear dust level detector. Each baghouse compartment has one blade type inlet damper, two outlet poppet type valves, and one reverse air poppet valve. All valves and dampers are pneumatically actuated by solenoids operated from the control panel.

The operation of the two baghouses is controlled independently by one Texas Instruments STI Programmable Controller utilizing a Control Junctions, Inc. multiplexer system. Reverse air cleaning can be initiated by baghouse differential pressure or be set to clean continuously with a variable time delay between cleaning cycles. One compartment per baghouse is cleaned at a time. One to three reverse air cycles can be programmed for each cleaning period. Reverse air is provided by drawing cleaned flue gas from the baghouse outlet via three (two operating, one stand-by) 200 hp reverse air fans.

For start-up and upset conditions, the baghouses can be bypassed through double louver dampers which connect the inlet and outlet flues. These dampers will open for emergency bypass on high or low inlet gas temperature or high differential pressure. During baghouse operation, reverse air is utilized to purge the area between the two sets of louvers.

START-UP OF THE AVS DRY SCRUBBER AND BAGHOUSE

Process and mechanical checkout of the various scrubber sub-systems and the baghouses began in November, 1982, and continued through May, 1983. On May 24, 1983, the first coal firing occurred.

Throughout initial firing of the boiler on oil and the first few days of initial coal fire, the baghouses were operated on bypass. During this bypass period, the compartment access doors were cracked to vent fresh air into the compartments and the hopper heaters were turned on to help preheat the baghouse.

On May 27, 1983, it was felt that coal firing was stable. The baghouse access doors were sealed and the bypass dampers closed, bringing the baghouse on line for the initial coating of flyash. The boiler was operating at approximately 60% load and the flue gas entering the baghouse was approximately 320°F. Unfortunately, within a few hours after the initial bag coating process had begun, the boiler control system suffered a momentary power loss, causing a unit trip. The baghouses were put back on bypass for reigniting of the boiler.

Approximately 24 hours later, the baghouses were brought back on line. The reintroduction of flue gas to the baghouses was at essentially the same load and temperature as the previous day. The initial pressure differential across both baghouse inlet to outlet flues was two inches water. After about six hours, the pressure differential had risen to four inches water, and the first reverse air cleaning cycle was initiated.

On June 8, 1983, after two weeks of bag coating with flyash only, SO_2 removal with single pass operation (lime only) in the scrubber commenced. The first scrubber operation with recycled material being added to the slurry occurred on June 11, 1983. On June 12, the boiler was taken off-line for a one month shutdown for turbine fine screen removal and boiler chemical cleaning and was back on-line on July 10. The scrubber operation was gradually brought to design operating conditions (eg. 35% total solids and 170°F spray dryer outlet temperature) during the next two months by increasing the solids content of the slurry with recycle material and decreasing the outlet temperature with higher slurry feed flow rates.

PROBLEMS ENCOUNTERED AND THEIR CURRENT STATUS

Dry Scrubber

The initial operation of the scrubber and baghouses followed the anticipated schedule and at no time delayed the boiler start-up or operation. The overall success of the start-up effort was probably the result of Joy/Niro's experience at the Riverside full scale demonstration plant. There were, however, initially some mechanical problems with various components of the absorber product conveying system, atomizer wear plates, and the atomizer head tanks.

The recycle lifting conveyor has horizontal sections at the upper and lower extremes. During the first weeks of recycle operation, a portion of the paddle type flights of the conveyor experienced fatigue failure due to the wearing of the paddle tips on the center plate in the lower 90° bend of the conveyor casing. To remedy this, a guide rail was added supporting the chain in the conveyor bend, preventing the paddle flights from riding on the center plate.

Dusting from the conveyors made it necessary to redesign shaft and dump gate seals, and to increase the venting of the recycle bin. Operation of the variable speed drag link conveyor which feeds ash from the recycle bin into the mix tank indicated that the 3 hp motor originally supplied had insufficient power at the lower range of the operating speed. A 5 hp drive will be installed.

The interior of the atomizer wheel is equipped with a ceramic bottom wear plate. During the early operation, several of these plates cracked, possibly due to thermal shock as the wheel came into contact with the flue gas and then the feed slurry. A procedural change, initiating atomizer wheel protection water flow simultaneously with initiation of chamber gas flow, has solved this problem.

The atomizer head tanks (1 foot diameter by 6 feet tall) are equipped with a vertically mounted, internal, cylindrical screen. The feed slurry pumped from the feed tank enters the bottom of the head tank and into the center of the screen. The portion of the slurry which is delivered to the rotary atomizer must pass through the cylindrical screen wall in order to enter the atomizer feed line. The remainder of the slurry continues through the center of the head tank and out the open top of the screen to the overflow pipe, which returns the slurry to the feed tank. The screen self-cleans as this return flow through the center of the screen washes it's inner surface. During the early scrubber operation at the design spraydown rate and full gas flow, the head tank screens periodically restricted flow to the atomizers. The problem was found to be insufficient flow through the interior of the screen to provide the self-cleaning effect. Using the adjustable pitch sheaves provided on the feed pumps, the pump speeds were increased, thereby providing additional head tank flow.

One of the prime factors for successful drying of the feed slurry is maintaining sufficiently high feed solids. A basic concept of general spray drying technology is that

It is easier to dry a thick, or higher solids slurry, than a thin, or low solids one. During the start-up operation of the scrubber, periodic tripping of the recycle feeder to the mix tank caused low feed solids. This, along with a period of operation without the implementation of the gas flow feed forward signal to the atomizer feed controllers, caused some intermittent spray dryer wall deposits. The deposits did not interrupt scrubber operation, however, pluggage of the spray dryer bottom discharge did occur as dry lumps fell from the dryer walls. An interlock of outlet temperature to the feed solids has been programmed into the controls to automatically raise the outlet temperature in the event of a low feed solids content, thereby reducing the possibility that deposits will develop within the dryers.

Baghouse

The first and perhaps most significant problem encountered with the baghouse operation was high differential pressure. From the time of the initial reverse air cleaning at four inches pressure differential, it became apparent that the reverse air was not effectively removing the dust cake on the bags. During the first two weeks of operation, baghouse pressure drops ranged from four inches to over nine inches water gage, depending on boiler load. For full load boiler operation, the baghouses had to be bypassed to assure that the I.D. fans would not enter a stall condition.

Although it was realized that the higher gas temperatures through the baghouses (since the spray dryers were still off line) were resulting in higher gas volumes and a lower reverse air fan efficiency, these factors could not explain the excessive pressure loss. Inspection of the fabric dust layers showed heavy build-ups of caked dust on the bags. Laboratory analysis of fabric samples confirmed that the bags had experienced a dew point encounter causing soluble sulfate formations resulting in agglomerations and nodula formation. The level of agglomeration and cake pluggage was not considered severe and the dust cake was easily removed by manual agitation of each individual bag.

Current baghouse pressure differential has stabilized at approximately five and one-half inches water gage at design conditions.

During the initial two weeks of operation, stack opacities of 10-20% were observed. The reason for visible emissions from the baghouse was apparent after the June 12, 1983, shutdown, when it was discovered that several of the originally supplied bypass damper blade seals had torn loose, possibly due to high boiler load bypass operation. Velocities through the louver damper were in excess of 100 ft/sec. The design of the blade seals was modified and new seals were installed.

COMPLIANCE AND PERFORMANCE TESTS

Compliance Requirements

The dry scrubber system is designed to remove 62% of the SO_2 when the average sulfur coal (0.68%) is burned and 78% for the maximum sulfur coal (1.22%). Unit 2 of the Antelope Valley Station is scheduled to begin commercial operation in July, 1986. At that time, Unit 1 SO_2 removal requirements will increase to 81% for average sulfur coal and 89% for maximum sulfur coal. The SO_2 emissions from Unit 1 and Unit 2 combined are limited to 3,845 pounds per hour (0.39 pounds per million Btu), as a requirement of the Permit to Construct issued by the North Dakota State Department of Health.

Compliance Tests

Stack emission tests were conducted on August 23, 1983, to determine compliance with the Permit to Construct. The spray dryers had only operated for a total of about five weeks at the time of the emission compliance tests. The scrubber system had not reached design operating conditions, but was determined to be capable of meeting emission compliance requirements, so the tests proceeded. The results of the compliance tests are summarized in Table 1.

Performance Tests

Performance tests to determine guarantee compliance with the Joy contract were conducted September 26-30, 1983. During the tests, the SO_2 and particulate removal system operated at the conditions specified in the contract for normal operating conditions with the average sulfur coal.

EPA reference methods were used to determine SO_2 and particulate concentrations and gas flows at the inlet to

Table 1. Emission Compliance Test Results

Parameter	Test Result	Permit Limitation
Gross Generation, MW	468	
Stack Gas Flow, ACFM	1,880,000	
Stack Temperature, °F	222	
Total Solids in Slurry, %	24	
Inlet SO_2, PPM	668	
Stack SO_2, PPM	171	
Stack SO_2, #/hr	2,308	3,845
Stack SO_2, #/MKB	.46	
SO_2 Removal, %	74	
Stack NO_x, #/hr	2,222	2,465
Stack NO_x, #/MKB	.45	
Stack Particulates, #/hr	72.7	210
Stack Particulates, #/MKB	.015	

and in the stack, and SO_2 and gas flow of the bypass reheat duct. Lime consumption was determined from the pebble lime weigh belt feeder, with the lime being analyzed for available CaO. The lime consumption was verified with the slaked lime slurry analysis and flow rate. Pressure drop was confirmed from static pressure readings taken at several locations to account for the pressure drop from the outlet of the air heater to the inlet of the ID fan, and from the outlet of the ID fan to the entrance of the stack.

An enthalpy balance was used to confirm the gas flows and the pressure drop used in the evaluation. Electrical energy consumption was measured with a watt transducer.

Samples of coal, raw lime, slaked lime, feed slurry, treated water, dilution water (cooling tower blowdown), scrubber inlet flyash, and recycle products were collected and analyzed. Temperatures were taken of feed slurry, and spray dryer inlets, outlets, bypass reheat, and stack gases.

Test Results

Preliminary results of the guarantee performance tests on the SO_2 and particulate collection system are listed in Table 2. As of this writing, sample analyses have not been completed to determine the actual stoichiometric ratio. An estimate is used here based on correlation of available data.

During the performance tests, the ash conveying system from the baghouse to the recycle bin was not conveying adequate quantities of ash. To replace that ash, a truck was used to transport ash from the ash storage silo to the recycle bin. Preliminary results indicate that this ash contained higher alkalinity than the ash collected from the baghouse. This issue is currently under investigation to determine its impact on stoichiometric ratio, if any.

Testing was also conducted under conditions without reheat by-pass to assess the guarantees at 62%, 81%, and 89% removal. The preliminary results indicate a slight increase in stoichiometric ratio when the reheat by-pass was not used for 62% removal. A stoichiometric ratio of about 0.9 was used for 81% removal, which was well below the guarantee of 1.1.

A test at the high removal rate (89% guarantee) was

Table 2. Performance Test Results

Parameter	Test Result	Guarantee Conditions
Spray Dryer Inlet Gas Flow, ACFM	2,039,641	
Spray Dryer Inlet, °F	314	
Spray Dryer Outlet, °F	166	
Approach to Saturation, °F	29	
Total Solids in Slurry, %	41	
Baghouse Outlet, °F	189	185
Inlet SO_2, ppm	655	800
Outlet SO_2, ppm	220	304
Outlet SO_2, lb/hr	2,973	3,844
SO_2 Removal, %	63.9	62
Stoichiometric Ratio (for 63.9% removal)	.47	
Stoichiometric Ratio (corrected for 62% removal)	.45	.54
Inlet Particulate, gr/dscf	2.8496	
Outlet Particulate, gr/dscf	.0025	.012
Outlet Particulate, lb/hr	24	210
Particulate Removal, %	99.91	
Total Pressure Drop, in. W.G.	14.96	13.4
Power Consumption, KW	2,341	2,757
Reheat By-pass, 10^6 Btu/hr	7.1	7.2

conducted using data obtained from the plant monitoring equipment, and correlated to the previous EPA reference method tests. A stoichiometric ratio of 1.45 is guaranteed for 89% removal with the maximum design sulfur content of 1,380 ppm SO_2 into the system. The SO_2 inlet during the test was only 442 ppm. Removal of 92% was obtained with a stoichiometric ratio of about 1.0.

FUTURE OPERATION

Although some preliminary optimization testing has been conducted on the scrubbing system, additional testing will resume when the unit returns on-line in December. The testing will focus on maximizing lime utilization and minimizing differential pressure across the baghouse and spray dryers. The parameters to be studied during this testing will be stoichiometric ratio, approach temperature, and reheat usage.

Further testing of this system will also be conducted at 81% and 89% SO_2 removal conditions. The testing at these conditions will be used to confirm operation for the stringent removal requirements when Unit 2 begins operation and when the maximum sulfur content coal is burned. Other tests will be performed to determine the actual turn-down capabilities of the spray dryers.

Engineering studies are currently planned to investigate the possibility of decreasing the system differential pressure by modifying existing duct work. Replacement of the existing slurry feed pumps and piping is also being considered in order to operate at lower approach temperatures. Lower approach temperature operation should improve lime utilization and assist in maintaining the plant water balance.

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CHARACTERIZATION OF AN INDUSTRIAL SPRAY DRYER AT ARGONNE NATIONAL LABORATORY

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ABSTRACT

Argonne National Laboratory (ANL) is operating an industrial-scale, coal-fired boiler with a flue gas cleaning (FGC) system consisting of a spray dryer and fabric filter. This paper presents a description of the FGC system together with a status report for an EPA-sponsored project being carried out by ANL to characterize the operation of the system. This project involves a design and economic analysis of the FGC system, determination of waste characteristics, and analysis of system operation through monitoring of inlet/outlet gas streams and sampling of various process streams. Preliminary data and material balances are presented in the paper, as well as a proposed performance model based on an analysis of key operating parameters.

INTRODUCTION

"Dry" scrubbers consisting of spray dryers and fabric filters or electrostatic precipitators have recently emerged as a viable pollution control option for coal-firing electric utilities and industries. Such integrated flue gas cleaning (FGC) systems have several potential advantages over conventional wet scrubbers, including higher reliability, lower cost (for combined sulfur oxides and particulate matter control), lower consumption of energy and water, and greater ease of waste disposal. The latter point is due primarily to the production of a dry powdery waste that is readily handled, transported, and disposed of in landfills.

Most existing or planned dry scrubbers have been designed for systems firing low- to medium-sulfur coals. However, Argonne National Laboratory (ANL) has been operating the first commercial spray-dryer FGC system designed for high (3.5%) sulfur coal since November 1981. This system treats the flue gas from a spreader-stoker boiler rated at 170,000 lb/h of steam. Sulfur dioxide (SO_2) removals of about 80% are achieved routinely to meet the emission limit of 1.2 lb/10^6 Btu mandated by the State of Illinois.

Because of the unique nature of this system, the extensive provisions made for monitoring and sampling, and the ready availability of research facilities at ANL, the U.S. Environmental Protection Agency* (EPA) is sponsoring a program for characterization of the FGC system. The objective of this program is to document design and construction experience with the ANL system and to achieve a thorough understanding of its operation so as to develop predictive performance correlations. The first part of the program utilized the information compiled during the specification, bidding, construction, and initial operation phases. The analysis of specifications revealed items critical to the procurement of a spray dryer FGC system. The bid analysis reviewed the range of bids received on the basis of both capital and operating costs. This analysis of equipment and installed costs will give potential users of spray dryer FGC systems an insight to expected capital and operating expenditures.

A second part of the program involves characterization of the FGC system over a 60 to 90 day period. This characterization will entail extensive data acquisition and stream sampling, and will relate system operation to process conditions.

A third program activity is determination of the properties of the dry waste produced by the FGC system. Various mixtures of flyash and spent sorbent have been subjected to the EPA Extraction Procedure (EP) Toxicity Test[1] in order to determine the extent and composition of any potential leachate. The results are useful in evaluating disposal characteristics and determining the partitioning of coal-derived trace elements between ash and sorbent particles.

This paper presents a brief description of the system, a summary of the EP test results, and selected operating data obtained in a preliminary test program. The derivation of material balances around key system components is demonstrated, and a simple performance model is proposed.

SYSTEM CHARACTERISTICS

ARGONNE BOILER NO. 5

The spray dryer system is installed on Boiler No. 5 at the ANL heating plant. No. 5 is the largest of the five boilers at the plant, with a nameplate rating of 170,000 lb/h of saturated steam at 200 psig. It is a Wickes

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*Emissions/Effluent Technology Branch, Industrial Environmental Research Laboratory, Research Triangle Park, N.C., Theodore G. Brna, Project Officer.

(now Combustion Engineering) spreader-stoker unit that was installed in 1965 with the capability to fire either coal or gas. Oil-firing capability was added as a backup in 1973 when coal use was discontinued at ANL.

In 1980, as part of the DOE effort to maximize coal use and as an economy measure for the Laboratory, the coal handling equipment at the heating plant was refurbished and Boiler No. 5 was reconverted to coal firing.[2] However, it was also necessary to install an FGC system capable of providing compliance with the applicable State of Illinois emission regulations shown in Table 1 since existing equipment consisted solely of cyclone separators for flyash collection. These continue to be used and considerably reduce the flyash burden on the FGC system.

Several midwestern and eastern coals have been successfully fired in Boiler No. 5 since the reconversion, ranging in sulfur content from about 1.5% to 4.5%. Table 2 gives a typical analysis for the coal currently being used, which is from the Illinois basin and is representative of the fuel for which the FGC system was designed. During acceptance tests using that coal, SO_2 emissions of 0.95 $lb/10^6$ Btu and total suspended particulate emissions of 0.007 $lb/10^6$ Btu were measured.

FLUE GAS CLEANING SYSTEM

Although spray dryer systems are commonly referred to as "dry," they do involve the preparation and handling of a slurry. For this discussion, it is convenient to divide the system into its wet and dry subsystems. The wet subsystem encompasses pebble lime storage, recycle powder storage, lime preparation, and slurry preparation. The dry subsystem is basically the gas train and associated equipment, and includes the spray dryer, fabric filter, ducts, booster fan, and air compressor. The relationship between major components is illustrated in Figure 1.

The feed slurry is composed of a mixture of fresh slaked lime and recycled "spent" sorbent. The recycled material contains a significant component of unreacted calcium and thus raises the effective Ca/S ratio above that indicated by the fresh lime alone. Pebble lime from the lime silo is fed to a weighbelt feeder, which supplies a paste-type slaker where the lime is converted from calcium oxide (CaO) to calcium hydroxide ($Ca(OH)_2$). Some water is added at the outlet of the slaker to dilute the milk of lime to a concentration of approximately 15% (by weight) solids. This milk of lime then flows to the milk of lime tank through a rotary screen, which removes the "grits," or inerts.

From the milk of lime tank, the lime slurry is pumped to the slurry mix tank. In this tank, milk of lime, recycled spent sorbent powder, and some additional water are combined into an approximately 35-40% (by weight) slurry. This slurry is then pumped to another rotary screen (to ensure

TABLE 1. EMISSION LIMITS FOR BOILER NO. 5

Pollutant		Limit
Sulfur Dioxide		1.2 lb/10^6 Btu
Particulate Matter		0.1 lb/10^6 Btu
	and	$\leq$20% Opacity

TABLE 2. TYPICAL COAL CHARACTERISTICS

Coal Parameter	Value
Heating Value, Btu/lb	12,027
Moisture, %	9.59
Ash, %	7.40
Carbon, %	66.98
Hydrogen, %	4.65
Nitrogen, %	1.48
Sulfur, %	3.32
Chlorine, %	0.06
Oxygen, %	6.52

removal of any large particles which might clog the atomizer nozzles) and then to the slurry feed tank. Overflow from the slurry feed tank goes back into the slurry mix tank, so there is a continual circulation between the tanks. Slurry from the feed tank is pumped, at a constant high flowrate, to a head tank located above the atomizer. A control valve regulates the amount of slurry fed to the atomizer, with the excess being returned to the feed tank. This returned slurry also passes through the rotary screen that is filtering the stream from the slurry mix tank.

Flue gas, exiting the boiler's induced draft (ID) fan, passes into a modified breeching at the existing stack. A guillotine damper diverts the flue gas flow into the FGC system ductwork leading to the spray dryer. This inlet ductwork splits the flue gas into two streams. One stream, with about 60% of the gas flow, is directed into a roof gas disperser, located on the top of the spray dryer (Figure 2). The remainder of the gas stream enters a central gas disperser, located in the middle of the spray dryer. Both gas

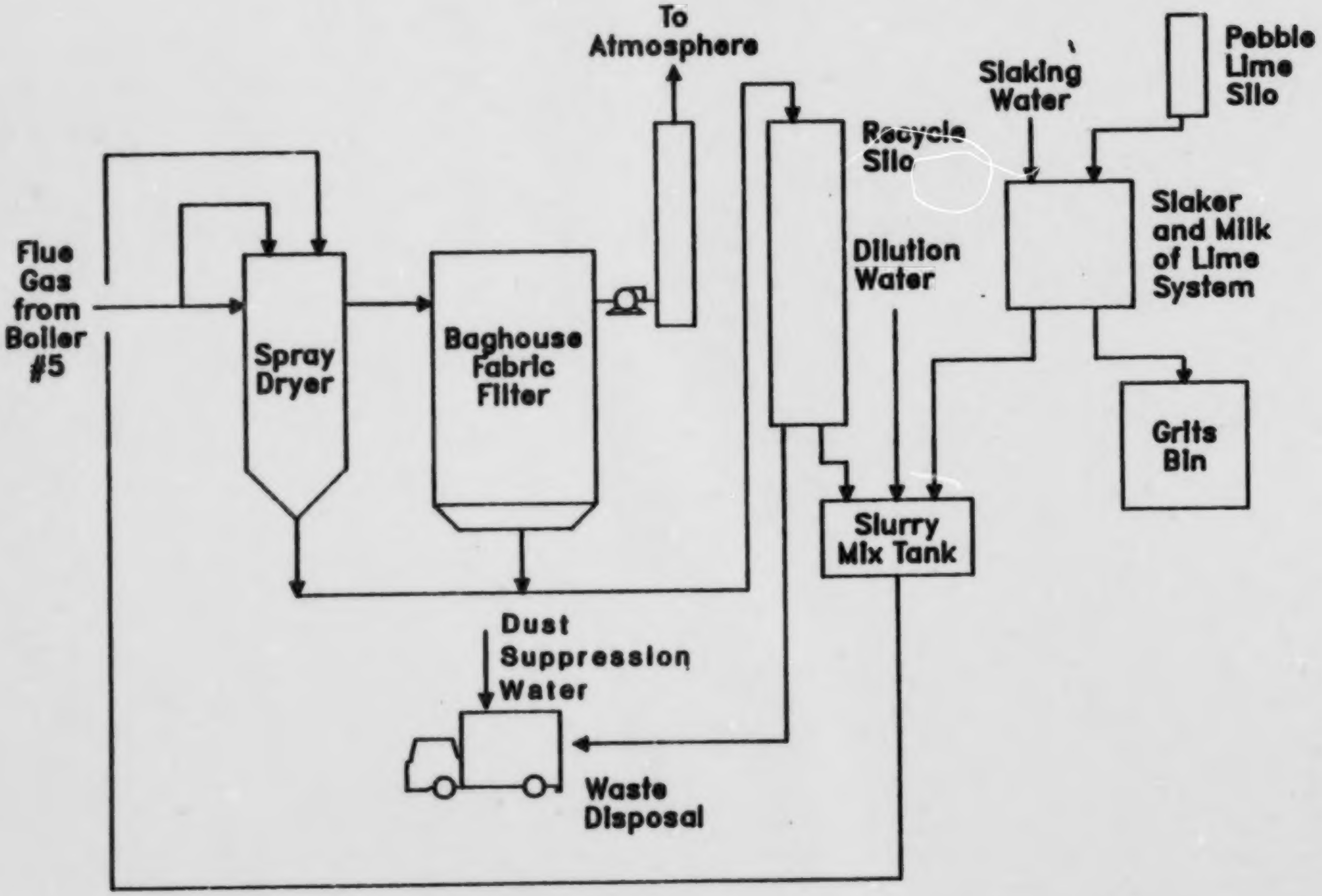

Figure 1. Argonne National Laboratory FGC System

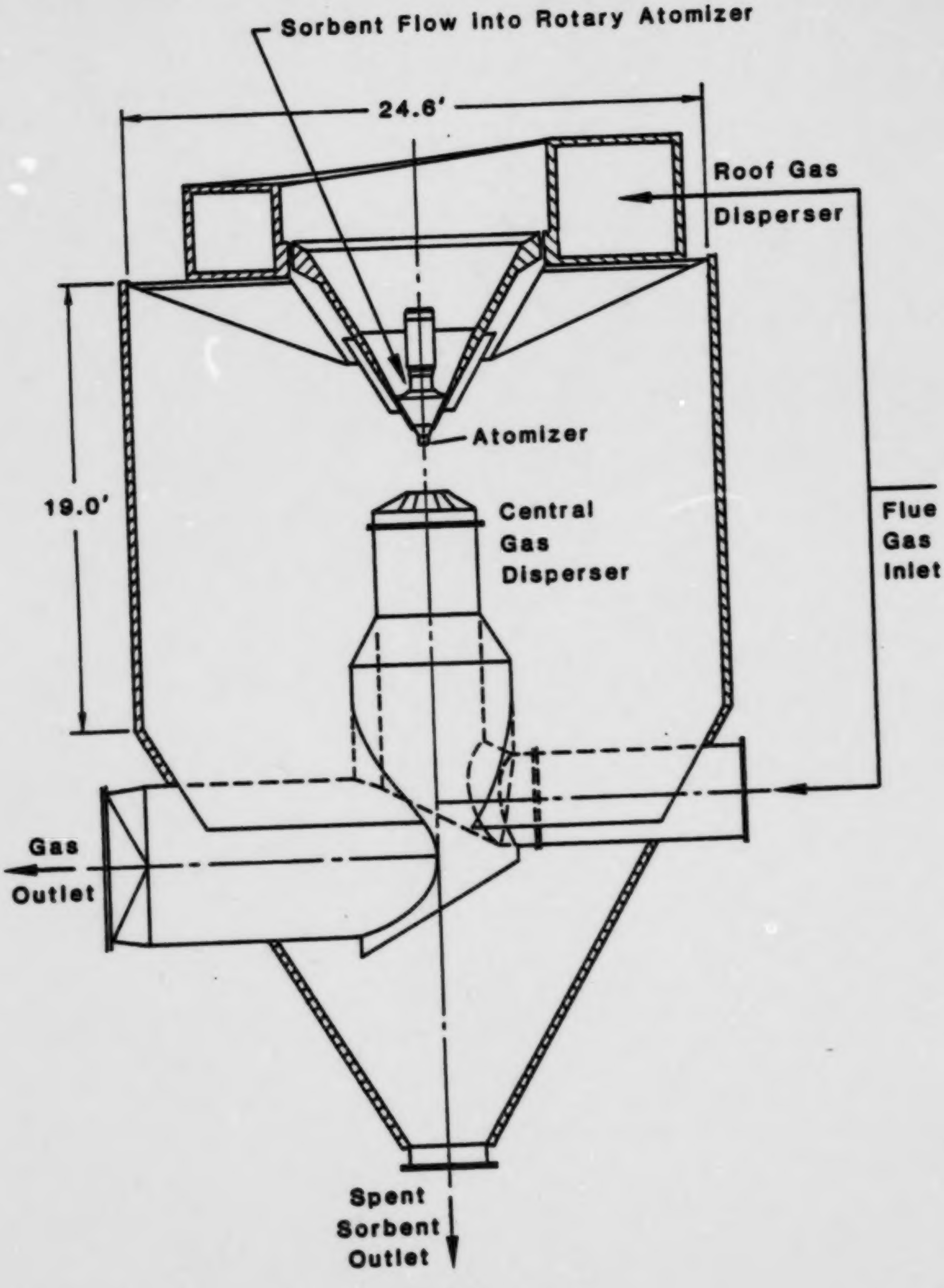

Figure 2. Spray Dryer

streams, upon entering the dryer, are given circular motions with their main directions of flow being opposed to each other. In the spray dryer the feed stream of slurry is atomized into fine droplets, through the centrifugal action of the rotating disk, and introduced into the gas stream. Some of the dry powder produced in the spray dryer falls to the bottom of unit and is picked up by a drag link conveyor for storage in the recycle/disposal silo.

The remainder of the powder is entrained in the gas stream and is carried into the four-compartment pulse-jet fabric filter. In the fabric filter these particles are separated from the gas stream by fiberglass bags. The cleaned flue gas exits the baghouse and travels through the outlet duct to the booster fan. From the booster fan the gas enters a new breeching section of the stack, and then into the stack itself for discharge to the atmosphere.

Wet Subsystem

Pebble Lime Silo--

This vessel is constructed of carbon steel, and is 12 ft in diameter with a 32 ft high straight side. There is a partial-cone lower section on the vessel, with a "live-bin" vibratory bottom. The silo is designed to contain up to 110 tons of high-calcium pebble lime, which is approximately a 7-day supply at full boiler load.

Recycle/Disposal Silo--

Standing 46 ft high with a 20 ft diameter, this silo is the larger of the two dry materials silos in the FGC system. As with the lime silo, this vessel is constructed of carbon steel. However, it has a flat bottom and is equipped with an air fluidizing system (air-slide) to assist in powder flow. Two outlets in the bottom are for recycle of spent sorbent and for sorbent disposal. The recycle silo has a capacity of 180 tons of dry powder which, at full boiler load, is sufficient for 7 days' accumulation.

Lime Slaking System--

This system consists of a Wallace and Tiernan paste-type lime slaker (Series A-758), with a capacity of 2000 lb/h, and a mechanical weighbelt feeder (Model 31-120AV). The feeder is equipped with an adjustable-speed belt drive and a transmitting switch for actuating the totalizing throughput counter. The slaker, made of heavy gauge steel, consists of a slaking compartment containing two sets of counter rotating intermeshing paddles for mixing, a dilution chamber with rakes for agitation, and a dust and vapor arrestor. In the slaker, water and pebble lime, in a ratio of approximately 2 to 1 by weight, are fed continuously into one end of the slaking compartment. Lime feedrate governs the operating rate of the entire system, and this feedrate is controlled on the basis of the liquid level in the milk of lime tank. Lime-paste consistency controls the addition of water. An increase in torque on the mixing shafts (indicating a thicker paste) opens a water-control valve to admit additional water.

Vibrating Screens--

These screens are both "Rotex Single Surface Liquatax Separators," 2 ft by 2 ft with 4 in. inlet nozzles and 3 in. outlet nozzles. One screen, the classifier screen, has a 40-mesh stainless steel screen and is at the outlet of the lime slaker. The purpose of this screen is to separate the "grits" from the slaked lime slurry. The overs from this screen flow through a pipe into a dumpster for disposal. The slurry passing through the screen discharges to the milk of lime collection tank.

The other screen, called the slurry feed screen, is a 6-mesh stainless steel screen which processes the slurry from the slurry mix tank as well as the returned slurry stream from the atomizer head tank. As with the classifier screen, the overs are sent to a dumpster for disposal. The slurry discharging from this stream falls into the slurry feed tank.

Tanks--

There are four tanks associated with the wet subsystem. These are the:

- Milk of Lime Collection Tank,
- Slurry Mix Tank,
- Slurry Feed Tank, and the
- Head Tank.

The milk of lime collection tank is an agitated, baffled tank with a 600-gal capacity. The tank, constructed from 1/4-in. thick ASTM-A283 plate, is 60 in. in diameter and 77 in. high. Based on the maximum milk of lime flow, the tank residence time is approximately 20 minutes.

The slurry mix tank and the slurry feed tank are identical in construction (920 gal), both being 60 in. in diameter and 101 in. high. As with the milk of lime collection tank, they are agitated, baffled tanks constructed of 1/4-in. A283 carbon steel. Their residence times are each approximately 20 minutes based on maximum slurry flows. In a sense these two tanks may be considered as a single large unit. This is due to the fact that the slurry mix tank continuously feeds the slurry feed tank at a rate of 45 gallons per minute (gpm), with the overflow from the slurry feed tank being returned to the slurry mix tank.

The head tank, located above the atomizer in the spray-dryer penthouse, is 6 in. in diameter and 4 ft high, and is constructed of carbon steel. It supplies a hydrostatic head for the slurry feed to the atomizer. Feed slurry is pumped to the head tank at a constant rate of 45 gpm. The flow of slurry to the atomizer from the head tank is modulated by a temperature-controlled valve in the line between the head tank and the atomizer. This valve responds to controller signals to ensure a constant gas temperature at the spray dryer outlet. Any slurry in the head tank that is not sent to the atomizer overflows into a return line to the slurry feed screen. The constant high flowrate to the head tank is necessary to ensure that, during periods of low slurry demand by the system, pipeline velocities are kept high enough (6-8 ft/s) to prevent the feed slurry from settling out in the long vertical feed line.

Pumps--

All pumps (milk of lime, slurry transfer, and slurry feed) are Warman 1- 1/2 in. x 1 in. Type BM with V-belt drives and overhead motors. These pumps are split-casing rubber-lined units with rubber-lined open impellors. The pumping systems are all redundant, with an identical "spare" pump connected with parallel piping and valves to the operating unit. Each pump is designed to deliver 45 gpm of fluid in order to ensure high velocities in the piping systems.

Agitators--

These Galigher Model GWO motor-driven agitators are, along with the pumps, the only rubber covered equipment in the entire sorbent preparation system. Their purpose is to ensure that all slurries stay in suspension, although the slurry mix tank agitator (which is fitted with two sets of impellors) has the additional duty of mixing milk of lime, recycled dry sorbent, and dilution water into a uniform slurry.

Dry Subsystem

Spray Dryer--

This is a dual-inlet spray dryer of carbon steel construction designed by Niro Atomizer. The unit is 27 ft 7 in. inside diameter, with a 19 ft straight section and a 19 ft long cone bottom. This cone bottom has a 3 ft diameter bottom outlet which is connected to a Niro-designed powder cooler, no longer utilized in the process. With an interior volume (less the central gas disperser volume) of 15,000 ft^3, the spray dryer has a gas residence time of 12 seconds at the maximum boiler load. This residence time is based on the inlet gas volume to the spray dryer. Should an average between the inlet and the outlet volumes be used, the value rises to almost 14 seconds.

The spray dryer was constructed with a dual gas inlet, such that 60% of the flue gas enters through the roof gas disperser, while the remaining gas flow enters via the central gas dispenser. These two gas streams are projected against each other, forming a turbulent swirl into which the slurry is injected by the high speed atomizer. This slurry mixes with the hot flue gas stream, reacting with the sulfur dioxide, and drying into a fine powder. Some of this powder falls to the bottom of the spray dryer, where it is picked up by a drag-link mechanical conveyor. The remainder of the powder is entrained in the gas stream and exits the spray dryer.

The atomizer used in the Argonne FGC system is a Niro Model F-160 equipped with a 200 hp 3570 rpm motor. A gear drive, within the atomizer body, increases the rotational speed to 14,800 rpm. This high speed spins the 8-1/4 in. diameter atomizing disk to a tip speed of 533 ft/s. The atomizing disk is constructed with eight equally spaced 3/8-in. diameter holes, through which the slurry is injected into the gas stream.

Baghouse Fabric Filter--

The baghouse fabric filter is a Joy Manufacturing, Model 6012, pulse-jet unit. It consists of four compartments, each having 28 rows of 10 filter bags. The filter compartments are constructed of carbon steel, and equipped with electrical plate-type heaters on the hopper (lower surface) walls. These hopper heaters are thermostatically controlled, and ensure that no moisture condenses in the compartments. Dust-laden gas enters each filter compartment in the lower plenum. The gas travels upward, through the bags, into the upper plenum of the filter compartment. From the filter compartment, the gas travels through ductwork to the booster fan.

The filter bags are fabricated of a woven 16 oz fiberglass fabric with a Teflon coating. These filter bags are 6 in. in diameter by 12 ft long and are fitted onto wire mesh cages. Each row of filter bags has a pipe, which is located over the bags and runs the entire length of the row. These pipes have holes in them above each bag and are connected through solenoid valves to a compressed air header. As each solenoid is actuated in turn, it sends a pulse of compressed air into the pipe. The compressed air exits through the holes above the bags, entrains some additional gas with the aid of a venturi nozzle fastened to the top of the bag cage combination, and travels down the length of the bag. This pulse of air breaks loose some of the cake collected on the outside of the filter bag. This broken cake falls to the bottom of the filter compartment, and into another drag-link conveyor.

Booster Fan--

This is a belt-driven induced draft fan manufactured by Chicago Blower Corporation (Model 5414). The fan wheel is 54-1/4 in. in diameter and is driven (via the belt drive) by a 250 hp, 1800 rpm motor. The fan is designed for 62,000 ft^3/min of gas at 19 in. W.C. static pressure while running at 1335 rpm. Flue gas from the baghouse fabric filter system is ducted to this fan and, from there, directly to the stack for discharge to the atmosphere.

RESULTS AND CONCLUSIONS

DESIGN AND ECONOMIC ANALYSIS

Five vendors submitted proposals in response to the Request for Proposals (RFP) issued by ANL for installation of a dry FGC system. In the spray dryer, four vendors proposed to use rotary centrifugal devices to atomize the lime slurry feed; the fifth proposed two-fluid nozzles to achieve the atomization. The diameter of the dryer chambers ranged from 23 to 27.6 ft, with straight-wall height ranging from 10 to 22 ft. Solids concentration of lime-slurry feed varied from 14 to 35% by weight. Estimated pressure drop across the dryer, at maximum boiler operating conditions, varied from 1.5 to 5 in. W.C., with estimated power consumption for operating the dryer varying from 66 to 111 kW.

For collection of flyash and spent sorbent, reverse-air and pulse-jet fabric filter baghouses were proposed, with Teflon-coated fiberglass bags used in four of the proposed systems. The gross air-to-cloth ratio for these baghouse systems ranged from 1.34 to 4.60, and the resultant net air-to-cloth ratio ranged from 1.81 to 6.16. The estimated pressure drop across the baghouse varied from 5.5 to 7.5 in. W.C., resulting in an estimated power requirement for operation of this subsystem of 17 to 30 hp.

As for overall system chemical and utility consumptions, the estimated lime (92% pure pebble lime) utilization rate for the proposed FGC systems ranged from 1,130 to 1,600 lb/h at maximum boiler capacity. Estimated total water usage varied from 20 to 40 gpm, and estimated total system pressure drop varied from 8.75 to 16 in. W.C., with a resultant estimated total power consumption for operation of these FGC systems ranging from 239 to 545 kW.

Both the total system capital costs and the estimated annual operating costs varied over a range of more than two to one among the various proposals. The final capital cost for the system as constructed was $3.5 million.

After completion of the vendor selection process, a letter of intent was awarded to Niro Atomizer, Inc. in the first week of November 1980. Ground was broken after general facility arrangement drawings were finalized for construction in the second week of April 1981, and erection of steel and platforms was completed on August 7, 1981. Delivery of major equipment commenced in the first week of June 1981 and by the first week of October, both the baghouse and spray dryer subsystems had been installed. All mechanical work, piping, instrumentation, and electrical work were completed in mid-October, and final painting and insulation were finished in the first week of November 1981. Construction was about 95% complete at this time (about 1 year after the start of the project), and Argonne and Niro decided to move into the start-up phase of the project.

WASTE CHARACTERIZATION

Waste from the Argonne system, which consists of a mixture of spent sorbent and flyash, is currently landfilled onsite. Spent sorbent is discharged (see Figure 1) from the recycle/waste silo into a dump truck. Approximately 12% of this material is flyash. The bulk of the flyash is removed from the flue gas by cyclones for reinjection into the boiler. That ash which is not reinjected is collected in a separate silo and transported by truck to the landfill site. Various mixtures of ash and spent sorbent have been subjected to the EP Toxicity Test[1] in order to determine the extent and composition of any potential leachate. The results indicate that the leachate composition is a function of both the pH of the leaching solution and the composition of the waste (ash/sorbent) blend.

Samples of both spent sorbent and flyash were collected under typical operating conditions while firing the coal specified in Table 2. The SO_2 removal was approximately 80% at the time. The spent sorbent (which also contains a small amount of flyash) was obtained from a sampling tap at the base of the storage silo, while the flyash sample was obtained from the boiler's cyclone separators.

The test procedures detailed in Ref. 1 were followed to generate liquid samples for elemental analysis. However, in several cases the pH could not be adjusted down to the target value of 5.0 ± 0.2 due to the high available alkali in the spent sorbent. If the pH was below 5.0 (pure flyash), no adjustments were made.

An Inductively Coupled Argon Plasma Spectrometer (ICAP) was used for elemental analyses of unblended flyash and sorbent samples. These measurements involved 32 different elements. For the leachate analyses, a Perkin-Elmer Model 5000 atomic absorption spectrometer was employed. Twenty-two elements were measured in these tests. In each case, two independent measurements were made for each sample.

One purpose of the analyses was to compare the waste streams from this system with those from other units in operation. Results for several elements are shown in Table 3, and are compared to three samples of spent sorbent from lime-based spray dryers treating flue gas from low-sulfur coal combustion, as analyzed by Radian for the Electric Power Research Institute (EPRI).[3]

Since the ANL boiler employs high efficiency cyclones for flyash collection, it is understandable that ANL's spent sorbent would contain less flyash than is found in the waste from a utility system. Thus, one would expect that the utility waste properties would be found between those of ANL's two waste streams if all other factors were equal. This fact is illustrated by the amounts of iron and aluminum (common constituents in

Table 3 ELEMENTAL ANALYSIS OF SAMPLES

Element	Concentration, ppm ANL Flyash	Research-Cottrell Waste[3]	Rockwell-Joliet Station Waste[3]	Joy/Niro Riverside Waste[3]	ANL Spent Sorbent
Aluminum	68,100	41,000	71,000	58,000	3200
Antimony	<500	<8	<8	<8	<500
Arsenic	<500	14.2	33	30	<500
Barium	130	400	190	350	20
Beryllium	40	5.1	8.5	4.3	4
Boron	500	ND[a]	ND	ND	500
Cadmium	70	<1.0	<1	<1.0	<5
Calcium	7100	170,000	97,000	150,000	283,000
Chromium	130	50	54	52	20
Cobalt	80	14	15	4.9	<10
Copper	150	84	81	16	40
Iron	367,000	22,000	44,000	20,000	6300
Lead	150	4.4	17	<20	130
Magnesium	3000	12,000	13,000	15,000	5500
Manganese	210	200	110	630	100
Mercury	ND	ND	ND	ND	ND
Molybdenum	<10	<0.5	8.7	16	ND
Nickel	460	26	39	215	30
Potassium	2600	3100	6300	4300	1700
Selenium	ND	6.1	4.7	<20	ND
Silver	<10	<0.5	0.5	<0.5	ND
Sodium	2800	12,000	15,000	2700	1400
Strontium	250	1800	1800	1900	170
Thallium	ND	<25	<30	<25	ND
Tin	<30	<30	<30	<36	<30
Titanium	4000	4900	5200	3100	230
Vanadium	270	70	120	580	30
Zinc	330	43	92	37	120

[a] Not Detected.

flyash) in the three utility sorbents compared to ANL's flyash and sorbent. Iron comprises over 35% of the total mass in Argonne's flyash, with aluminum being about 7%. In the spent sorbent, they are found to be 0.6% and 0.3%, respectively. The utility spent sorbents averaged 2.9% iron and 5.7% aluminum. Examining many of the trace elements normally found in flyashes, similar comparisons could be made between the ANL and utility wastes.

On the other hand, different relationships arise because the utility sorbents are all the result of combustion of low-sulfur western coal whereas ANL's flyash is from high-sulfur midwestern coal combustion. For example, the presence in the utility wastes of large amounts (compared to either the ANL sorbent or flyash) of potassium and sodium is probably due to basic differences between eastern and western flyashes. The fact that magnesium is detected in the utility samples at levels ranging from three to five times that found in the Argonne samples could be due to two factors. Certain western flyashes contain large amounts of magnesium compared to eastern ashes. Another factor could be related to the type or quality of lime used in the FGC system. Argonne uses a lime which contains approximately 2% magnesium, and it is possible that the lime used in the utility units had a higher magnesium content. Coal-related differences in the nature of the wastes are also seen in the analyses of barium and titanium (more prevalent in the utility sorbents than in either the flyash or sorbent from ANL) and boron (undetectable in the utility sorbents, but found at the same (high) level in both the ANL sorbent and flyash).

Figures 3-5 show typical results from analysis of leachate from the ANL sorbent/flyash mixtures. A more complete presentation may be found in Ref. 4. In general, three different pH-sensitive trends were observed: 1) metal concentration increasing with increasing pH, 2) a decrease in concentration with increasing pH, and 3) specific sensitivity (high concentration) at approximately pH 5.

The variation of boron concentration with percent sorbent (and hence pH) is given in Figure 3. In this case, concentration rises with increasing pH, although a slight dip is noticed at 75% sorbent, or about pH 11.8. Since wastes from a dry scrubber system will tend to be basic in nature when initially placed in a landfill or other site, it may be concluded that until the nature of the waste is modified by soil conditions extensive leaching of boron and other similarly behaving elements may occur.

Plant growth experiments using soil treated with dry scrubber wastes (0 to 4% by weight) have been conducted at Argonne.[5] For both corn and soybeans, the biomass decreased as the amount of dry scrubber wastes increased. Boron concentrations in the leaves of the corn and soybeans grown in soil treated with scrubber wastes were found to be up to 20 times greater than those found in the leaves of the control (no dry scrubber wastes).[6]

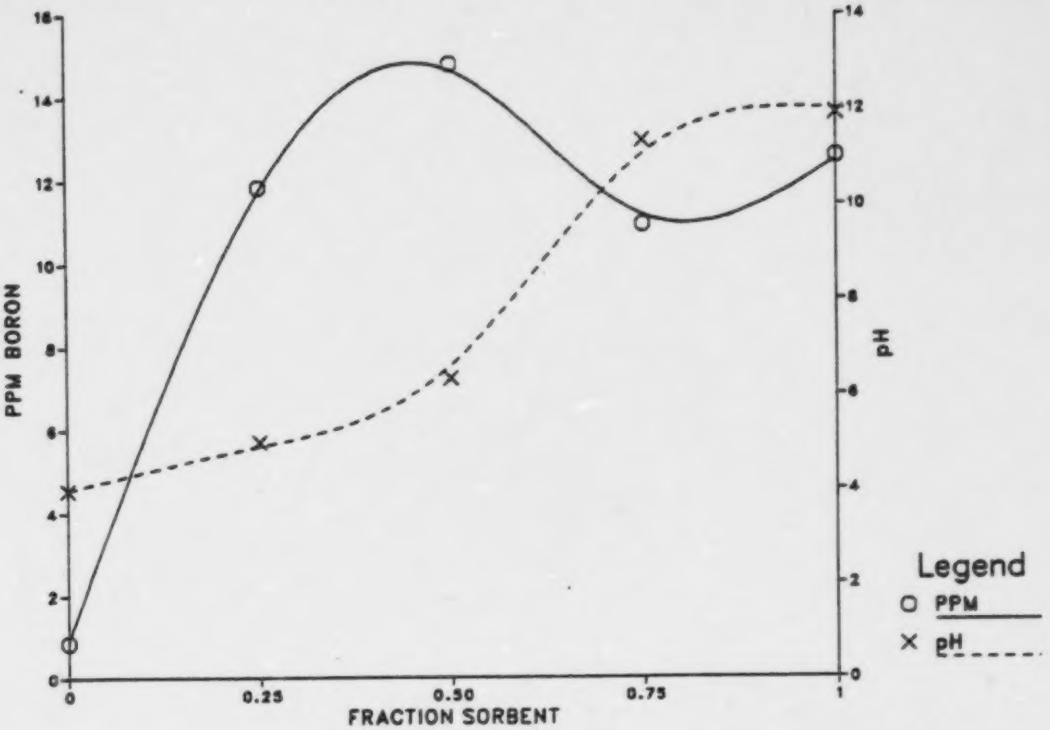

Figure 3. Boron Concentration in Leachate Varies with Fraction Sorbent and pH

Several elements, including cadmium (Figure 4), were found to have lower concentrations in leachate from the sorbent and sorbent mixtures than from the 100% flyash samples. This trend also follows for pH, in that the concentrations in the leachate decreased as pH increased. This effect is not unexpected, as it has been seen for these elements in many other wastes.[7] If leaching of cadmium compounds is actually retarded by high pH, impoundment of certain chemical wastes, which contain these and like materials, together with dry scrubber wastes may offer distinct advantages which should be explored.

Perhaps most interesting are the cases where response was greatest at a pH of approximately 5. Antimony (Figure 5) showed a concentration of approximately 60 ppb at pH 5 while levels for the 100% flyash (pH 4) and the pure waste sorbent (pH 12) were virtually nondetectable. Similar results were found for arsenic, mercury, and selenium. In the toxicity test procedure, pH control of the material is carefully specified. The reasons for the high leaching sensitivity of certain metals at, or about, the pH specified in the EP test procedure are not clear. Further work to determine the

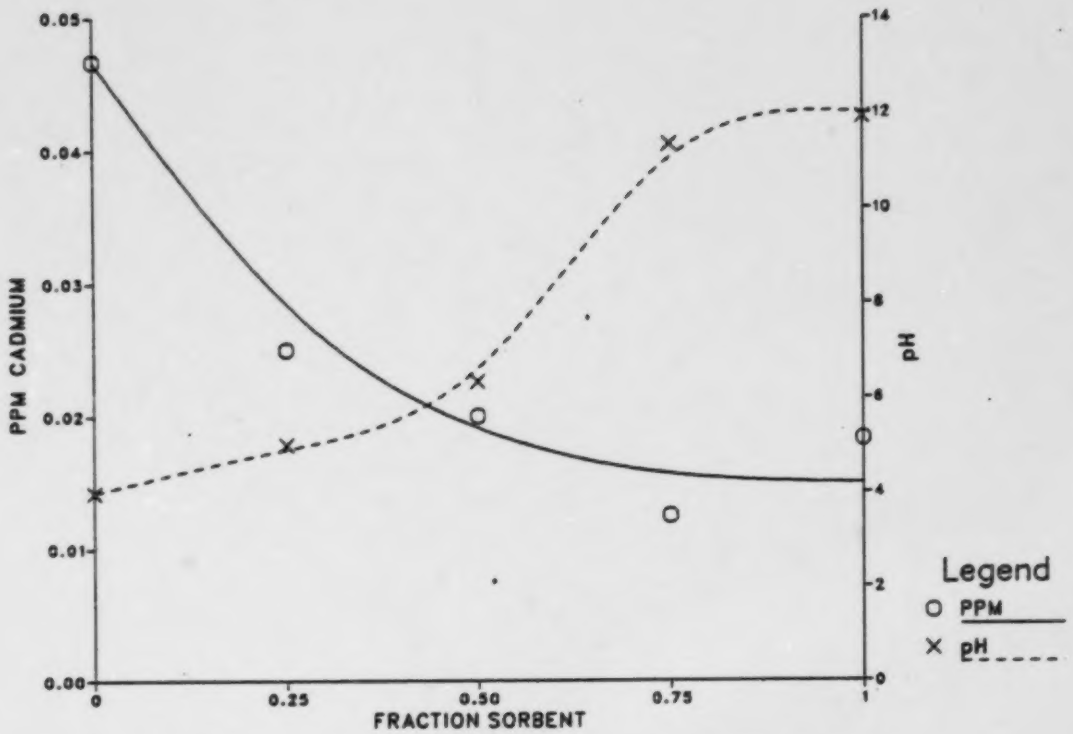

Figure 4. Cadmium Concentration in Leachate Varies with Fraction Sorbent and pH

reasons for this occurrence (e.g., the exact chemical form of the metallic compounds in the waste, and equilibrium conditions during the test) appears to be warranted.

Under RCRA, if the EP filtrate has a concentration of a substance that exceeds 100 times the Interim Primary Drinking Water Standard, then the waste can be classified as hazardous. Presently wastes from flue gas cleaning systems are excluded from regulation as hazardous wastes. Although no element had a concentration greater than 100 times the standard in ANL tests, it should be noted that the chromium levels were within 2% of the maximum.

HIGH SULFUR COAL TEST BURN

During April 1983, ANL and the Consolidation Coal Company conducted a test burn of very high sulfur (4.2-4.5%) coal in Boiler No. 5. This test, complementary in concept to the EPA-sponsored work, was undertaken to demonstrate:

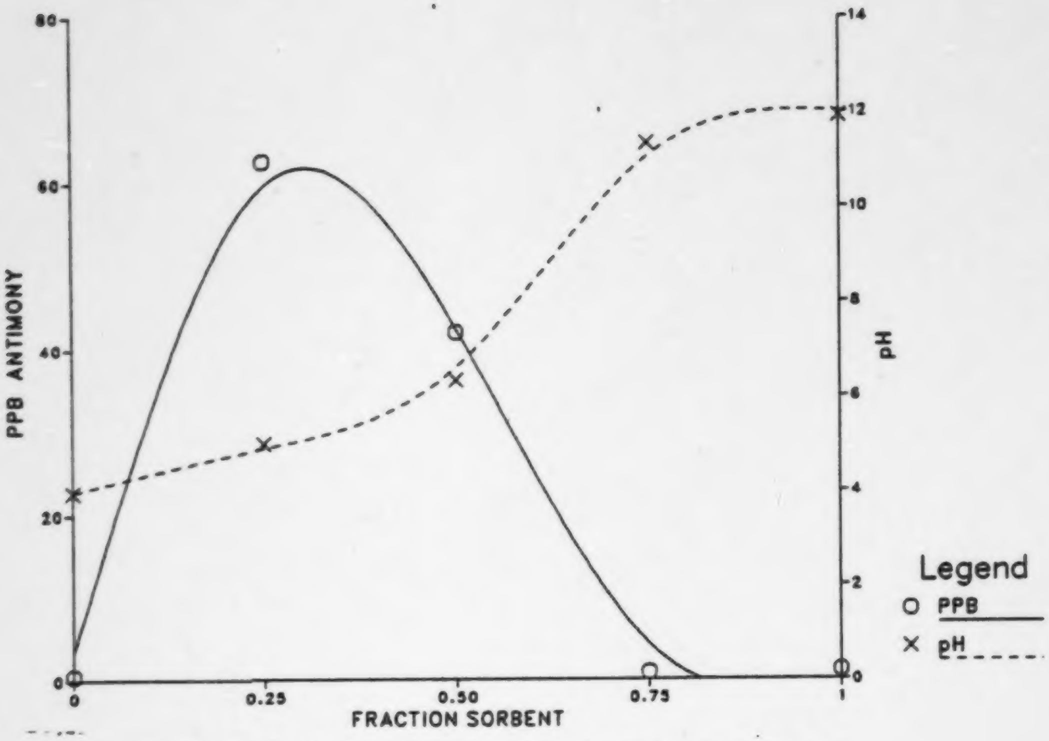

Figure 5. Antimony Concentration in Leachate Varies with Fraction Sorbent and pH

- Ninety percent or greater SO_2 removal over a 100 hour period,
- Ninety percent SO_2 removal at a stoichiometry (inlet moles of CaO divided by inlet moles SO_2) of 1.4 or less, and
- Seventy percent SO_2 removal across the spray dryer alone (the balance being attained in the baghouse).

The test consisted of extensive sampling and analysis of process streams, coupled with acquisition of process data from plant instrumentation and manual sampling (EPA Method 6) of the flue gas streams (inlet, exit from the spray dryer, and stack).

Although much of the data still remains to be analyzed, preliminary findings indicate that all three goals were achieved. In addition, a range of removals was achieved, which has allowed a comparison of the test results with data obtained using the nominal coal of 3-3.5% sulfur. A plot of percent sulfur removal as a function of external stoichiometry is given in Figure 6 and shows that, while 90% removal was attained for the 3.5% sulfur coal at a stoichiometry of about 1.15, a stoichiometry of approximately 1.3 was required for the same removal when treating flue gas from a 4.5% sulfur coal. It is important to point out that the stoichiometries being quoted are "external" stoichiometries and reflect the amounts of fresh pebble lime entering the system. From a more fundamental standpoint, it is important to examine the internal stoichiometry as a function of system performance. This is due to the fact that the internal stoichiometry also includes the available alkali in the recycle powder. It is this combined alkali which actually drives the SO_2 removal reaction in the spray dryer, and it should be taken into account when analyzing system operation. Based on data analyzed to date, Argonne has determined that the internal stoichiometry is (depending upon system operation) 15-20% greater than the external stoichiometry.

Frequent process stream samples were taken and analyzed during the testing period. For a set of samples taken on April 8, the compositions are shown in Table 4. At that time, the average SO_2 removal was 79.6%, with a temperature drop in the spray dryer of 195°F, and an approach (to dewpoint) temperature of 22°F. Examination of the results indicates that the ratio of calcium sulfite to calcium sulfate is on the order of 9 or 10 to 1. This is significantly different from wet lime scrubbing systems where ratios of approximately 3 or 4 to 1 are typical. It is felt that this difference exists due to the relatively short time (2-4 seconds) that a liquid droplet actually exists in a spray dryer. These short droplet lifetimes, in a dry scrubber, do not allow the oxidation reaction to proceed to the extent that it does in a wet scrubber, where droplet lifetimes are an order of magnitude longer.

The average process flows are shown in Table 5. These flows, together with the compositions in Table 4, allow acceptable material balances to be developed for system components. For example, taking a CaO material balance across the milk of lime tank we have:

a) In: $(860.2 \text{ lb/h})(0.946 \text{ available alkali}) = 813.75 \frac{\text{lb CaO}}{\text{h}}$

b) Out: $(8.7 \text{ gpm})(1.09 \text{ s.g.})[0.229 \text{ fraction } Ca(OH)_2](500 \frac{\text{lb/h}}{\text{gpm}})$

$$= 1085.8 \frac{\text{lb } Ca(OH)_2}{\text{h}}$$

$$= 821.7 \frac{\text{lb CaO}}{\text{h}}$$

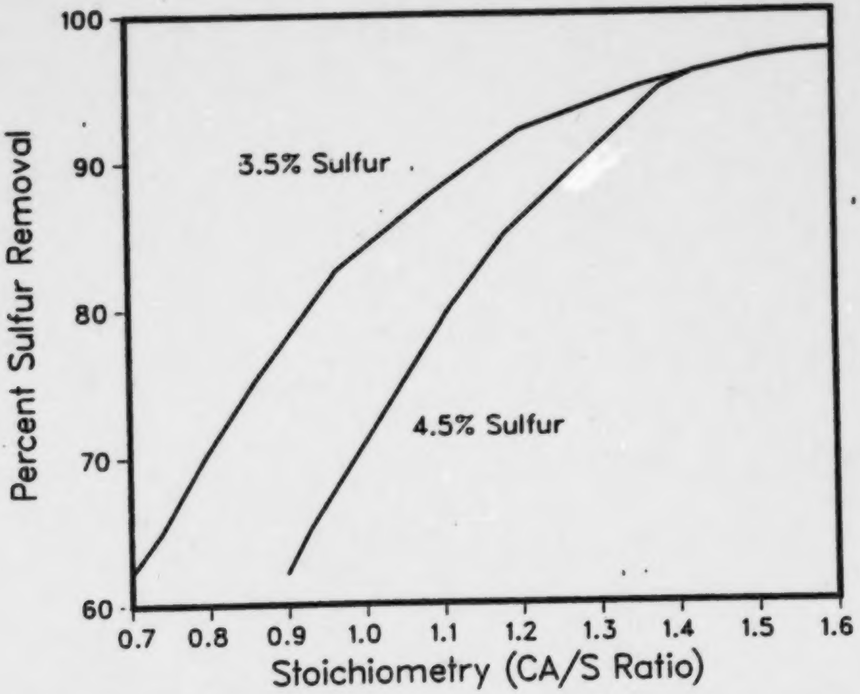

Figure 6. Percent Sulfur in Coal Affects Performance

TABLE 4. PROCESS STREAM COMPOSITIONS

Component	Percent by Weight Feed Slurry	Recycle Powder	Spray Powder	Milk of Lime
$Ca(OH)_2$	14.14	11.50	11.89	22.9
$CaSO_3 \cdot \frac{1}{2} H_2O$	9.60	60.87	54.42	-
$CaSO_4 \cdot \frac{1}{2} H_2O$	0.23	6.71	5.46	-
$CaCO_3$	1.04	5.98	8.46	-
H_2O	71.70	2.46	3.25	77.1
Inerts	3.29	12.48	16.52	-
	100	100	100	100

TABLE 5. AVERAGE PROCESS STREAM FLOWS

Pebble Lime[a], lb/h	860.2
Milk of Lime to Mix Tank, gpm[b]	8.7
Slurry to Spray Dryer, gpm[c]	14.77
Dilution Water to Mix Tank, gpm	4.6
Recycle Powder to Mix Tank, lb/h	1904
Spray Dryer Bottoms Rate, lb/h[d]	979
Waste to Disposal, lb/h[e]	3012
Stack Gas Flow, dscfm[f]	38,100
SO_2 in, lb/h	872
SO_2 out, lb/h	178

[a] @ 94.6% available alkali

[b] +8% level change in milk of lime storage tank over 24 hour period = 59 gallons

[c] No change in slurry tank level

[d] Based on 2.5 minute sample collected

[e] Includes 1.3 ft drop in silo level = 14,294 lb or 596 lb/h. Net disposal = 1814 lb/h based on a settled density of 70 lb/ft^3 and a solids moisture in the truck of 25%.

[f] Dry standard cubic feet per minute.

c) $$\text{Accumulation} = \frac{59\ \text{gal}}{24\ \text{h}}\ (1.09)\ (8.34\ \text{lb/gal})\ (0.229)$$

$$= 5.12\ \frac{\text{lb Ca(OH)}_2}{\text{h}} = 3.9\ \frac{\text{lb CaO}}{\text{h}}$$

d) In = Out plus Accumulation

$813.75 = 821.7 + 3.9 = 825.6$

e) Checking for error in the balance,

$$\Delta = \frac{825.6 - 813.75}{825.6} = 0.014, \text{ or } 1.4\%$$

Examining the operation of the system, it was determined that the same fractional removal of SO_2 is generally achieved across both the spray dryer and baghouse. For example, if a 50% removal of SO_2 is detected based on inlet and outlet concentrations at the spray dryer, then a 50% removal of SO_2 will be seen across the fabric filter, based on the inlet concentration to the filter.

Therefore, for the overall removal of SO_2 of 79.6% on the 8th of April, approximately 55% removal of SO_2 occurred in the spray dryer. Using this value in a balance around the dryer, it was found that for the lime reacting (9.79 lb moles/h), the products were:

$CaSO_3 \cdot \frac{1}{2} H_2O$ = 66%

$CaCO_3$ = 24%

$CaSO_4 \cdot \frac{1}{2} H_2O$ = 10%

which leads to the conclusion that carbon dioxide (CO_2) pickup does occur in the spray dryer.

Performing a similar balance over the baghouse show that:

$CaSO_3 \cdot \frac{1}{2} H_2O$ = 63.98%

$CaSO_4 \cdot \frac{1}{2} H_2O$ = 7.47%

$CaCO_3$ = 7.23%

$Ca(OH)_2$ = 4.66%

H_2O = 2.78%

Inerts = 14.17%

This gives the approximate composition of the powder exiting the fabric filter. It is clear that the system was not at steady state, since the combined compositions of the spray dryer powder and the baghouse (calculated) powder, does not equal the composition of the recycle powder. The attainment of steady state is complicated by the residence time of waste product in the recycle silo. This storage volume results in powder not appearing for recycle until several days following production (assuming plug flow in the silo).

PERFORMANCE MODEL

Argonne has identified what are believed to be the critical parameters for SO_2 removal in a spray dryer. These are:

a) $$\left(\frac{\text{Available Alkali}}{\text{Inlet SO}_2}\right)_{\text{SDA}} = (\text{SR})$$

This is the stoichiometric ratio (SR) based on the slurry fed to the spray dryer absorber (SDA).

b) $$\left(\frac{\text{Mass Available Alkali}}{\text{Total Mass of Solids}}\right)_{\text{SDA}} = (\text{RR})$$

This recycle ratio (RR) of available alkali injected into the spray dryer to the total mass of injected solids reflects the effect of spent sorbent recycle on system performance. The available alkali is determined by a chemical method (titration) that will detect the alkali whether it is on the surface of the particle or in the interior. If recycling of spent sorbent results in a partial "blinding" of available alkali by other compounds in the recycled powder, then the stoichiometry needed for a particular removal will be higher for a system with recycle than for one in which the operation is on a "straight-through" basis. An analysis of data and samples gathered in this program will allow for a determination of the significance of sorbent recycle on system operation.

c) $$[1 - \exp(-t/\tau)]$$

This factor is a reflection of the effects of boiler load (and hence gas residence time in the spray dryer) as well as the drying phenomenon for particles in the spray dryer. The variable, t, represents the gas residence time in the spray dryer, as an average of inlet and outlet gas volumes. The time constant, τ, is the time for drying an average sized particle in the spray dryer. This average drying time is thought to be a function of several factors including the percent solids in the slurry, the gas temperature drop in the spray dryer, and the average droplet size.

d) $$\left(\frac{T_{\text{Gas in}} - T_{\text{Gas out}}}{T_{\text{Gas in}} - T_{\text{Gas Sat'd}}}\right) = \frac{T_1}{T_2}$$

Many researchers have noted that the closer the exit gas temperature is to the dew point, or saturation temperature, then the lower the stoichiometry for a given removal. If the SO_2 removal reaction takes place within the droplets as they dry, then the temperature gradient across the spray dryer should play a key role. The numerator of the factor is the temperature drop of the flue gas across the spray dryer due to the evaporation of the slurry. The denominator is the maximum temperature drop that could be taken in the spray dryer before a wet product is produced. Another way of viewing this factor is that it is the ratio between the operation of a dry scrubber and a wet scrubber, where the gas is quenched to its dew point.

Combining the terms;

$$\text{Fraction } SO_2 \text{ removal} = (SR)^a \, (RR)^b \left(\frac{T_1}{T_2}\right)^c \, (1 - \exp(-t/\tau))$$

where a, b, and c are exponents which may, or may not, be equal to 1. Data obtained in the forthcoming performance tests on the FGC system will be used to check and calibrate this relationship.

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NOTES

1. Company Names and Products.

The mention of company names or products is not to be considered an endorsement or recommendation for use by the U.S. Environmental Protection Agency.

2. Units of Measure.

EPA policy is to express all measurements in Agency documents in metric units. When implementing this practice will result in undue cost or difficulty in clarity, IERL-RTP provides conversion factors for the non-metric units. Generally, this paper uses British units of measure.

The following equivalents can be used for conversion to the Metric system:

British	Metric
5/9 (°F-32)	°C
1 Btu/kWh	1055.056 J/kWh
1 ft 0.3048 m	
1 ft^2	0.0929 m^2
1 ft^3	0.0283 m^3
1 ft^3/min	4.719E-04 m^3/s
1 gallon per minute (gpm)	6.309 E-05 m^3/s
1 gallon (gal)	0.003785 m^3
1 grain	0.0648 gram
1 horsepower (hp)	746 watts
1 in. W.C.	249.08 Pascals
1 lb (avoir.)	0.4536 kg
1 lb/10^6 Btu	429.6 ng/J
1 lb/ft^3	16.01 Kg/m^3
1 ton (long)	1.0160 m tons
1 ton (short)	0.9072 m tons

UNPRESENTED PAPERS

AN ECONOMIC EVALUATION OF LIMESTONE DOUBLE ALKALI FLUE GAS DESULFURIZATION SYSTEMS

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ABSTRACT

Considerable work was done at the EPA Scholz plant facility in defining the process parameters for limestone double alkali flue gas desulfurization systems. In general this study work proved the viability of the process but uncovered several less than optimum operating parameters that needed further work. FMC and others have continued to work with the process and have defined operating parameter changes necessary to make the system commercially viable.

Limestone double alkali is especially appropriate for FGD systems applied to boilers burning relatively high sulfur (2% and greater) fuel. A discussion of site-specific design criteria which impact on the selection of FGD technology is included with a definition of the optimum parameters for the applications of limestone double alkali.

An in depth economic analysis of the system is included with comparisons to conventional limestone scrubbing technology. Cost comparisons are made by subsystems such as absorber system, reagent handling, storage and preparation system, solids waste production and disposal. The economic data presented is primarily a result of study work done by TVA under contract from the Environmental Protection Agency.

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I. INTRODUCTION

For the past five years there have been extensive efforts to develop a sodium-based double alkali process in which limestone, rather than lime, is used to regenerate the absorbent, thus retaining the advantages of using a solution as the absorbent liquid while avoiding the higher operating costs inherent with the use of lime. FMC Corporation, Combustion Equipment Associates, Inc. (now Thyssen - CEA Environmental Systems, Inc.--TESI), A. D. Little, Inc., Accurex Corporation, and the U.S. Environmental Protection Agency have been active in the development of a limestone double alkali process. These efforts have advanced the technology of the process to the point that a commercially acceptable limestone double alkali process seems assured.

Among the studies of limestone double alkali processes in the past few years, there has been an EPA-sponsored prototype scale evaluation at the Gulf Power Company's Scholz steam plant, an economic evaluation by the Tennessee Valley Authority, and an economic evaluation by the Stearns-Roger Engineering Corporation. This paper summarizes these three evaluations and discusses studies and test work by the FMC Corporation to develop a limestone double alkali process.

II. TESI PLANT SCHOLZ WORK *

Thyssen-CEA Environmental Systems, in conjunction with Arthur D. Little, conducted an EPA-sponsored test program at Gulf Power Company's Scholz steam plant near Sneads, Florida. This involved the conversion of a 20-megawatt lime based dual alkali system to operation with limestone regeneration.

Budgetary problems caused a reduction in the test program from six months to two months. However, even during the shortened program the process proved to be technically feasible.

A. SYSTEM DESIGN BASIS

The limestone dual alkali system at Scholz was based on firing the boiler with coal containing 2.9 to 3.4 weight percent sulfur. These sulfur levels corresponded to approximately 1900 to 2300 ppm SO_2 in the boiler flue gas. Since it was desirable to run tests at the high end of this inlet loading and to include even higher SO_2 inlet concentrations, an SO_2 injection system was provided. This allowed operation with both a normal and maximum load condition with regard to SO_2. Normal system operation was based on a gas flow of 40,000 dscfm with an inlet SO_2 concentration of 2200 ppm dry. The maximum condition was based on a 45,000 dscfm gas flow with an SO_2 concentration of 2650 ppm dry. Table 1 and Table 2 give additional details for the normal and maximum design cases.

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* Valencia, J.A., et. al., Evaluation of the Limestone Dual Alkali Prototype System at Plant Scholz, Report to EPA, Contract 68-02-3128, August 1981. (The majority of the data contained in the section is taken from this report.)

TABLE 1. DESIGN BASIS - NORMAL OPERATION

Inlet Gas:	
Flow Rate	40,000 scfm (dry)
SO_2	14.9 lbs/min (2,200 ppm dry)
Removal Efficiency:	
SO_2	95%
Absorber/Scrubber Feed:	
Absorber Tray Feed	2.4 gallons/1000 acf
Scrubber Recycle	16 gallons/1000 acf
Soda Ash and Limestone Feed:	
Soda Ash Feed Rate	0.07 mole Na^+/mole SO_2
Limestone Purity	95 wt.%
Limestone Feed Rate	0.986 mole available $CaCO_3$/mole SO_2
Waste Solids:	
Wash Ratio	4 displacement washes
Insoluble Solids	55 wt.%

TABLE 2. DESIGN BASIS - MAXIMUM LOAD OPERATION

Inlet Gas:	
Flow Rate	45,000 scfm (dry)
SO_2	20.2 lbs/min (2,650 ppm dry)
Removal Efficiency:	
SO_2	95%
Absorber/Scrubber Feed:	
Absorber Tray Feed	2.6 gallons/1000 acf
Scrubber Recycle	15 gallons/1000 acf
Soda Ash and Limestone Feed:	
Soda Ash Feed Rate	0.081 mole Na^+/mole SO_2
Limestone Purity	90 wt.%
Limestone Feed Rate	1.010 mole available $CaCO_3$/mole SO_2
Waste Solids:	
Wash Ratio	4 displacement washes
Insoluble Solids	55 wt.%

B. DESCRIPTION OF THE SCHOLZ SYSTEM

The test plant facilities at Scholz consisted of three separate areas: the boiler, the limestone dual alkali scrubbing system, and the waste disposal facilities, that were provided as part of a parallel EPRI sponsored program to evaluate landfill disposal of the waste cake generated. The waste disposal facilities will not be discussed in this paper.

1. Boiler

The dual alkali scrubbing system was installed on Unit No. 1 at Scholz steam plant, a 40 MW capacity (47 MW peak capacity) B&W pulverized coal fired boiler. An electrostatic precipitator designed for 99.5% particulate removal follows the boiler. A portion of the flue gas, equivalent to a 20-MW boiler load, was directed to the limestone dual alkali system. The remaining gas was exhausted to the main stack.

2. Limestone Dual Alkali System

During 1975 and 1976 the Southern Company and EPA sponsored a jointly funded test program, conducted by TESI, utilizing a lime dual alkali system installed at Plant Scholz. The limestone dual alkali system utilized during the TESI testwork was a modification of that plant. The modified dual alkali system consisted of four basic sections: absorption; regeneration; waste solids dewatering; and reagent storage and preparation. A process flow diagram is shown in Figure 1.

a. Absorption--

The absorption system consisted of a plumb-bob type venturi scrubber followed by an absorption tower. The variable throat venturi was designed for both particulate removal and SO_2 absorption. Typically, a utility type dual alkali system would take flue gas from a particulate removal device (electrostatic precipitator or fabric filter) and would not require a venturi scrubber. However, rather than removing the venturi, it was used to quench and saturate the flue gases.

The absorption tower was designed to be used either as a spray tower or a tray tower. The limestone dual alkali testing was conducted utilizing the tower with two trays.

b. Regeneration Section--

Spent recirculated scrubbing solution was bled from the absorber to a multi-stage reactor system. The reactor system consisted of five separate tanks--an existing 750 gallon primary reactor followed by four new 3400 gallon tanks in series to allow a total of approximately 100 minutes retention time. All tanks were cylindrical, baffled vessels with center mounted agitators. Tests were run with all five reactors in series with dry pulverized limestone fed to the first reactor.

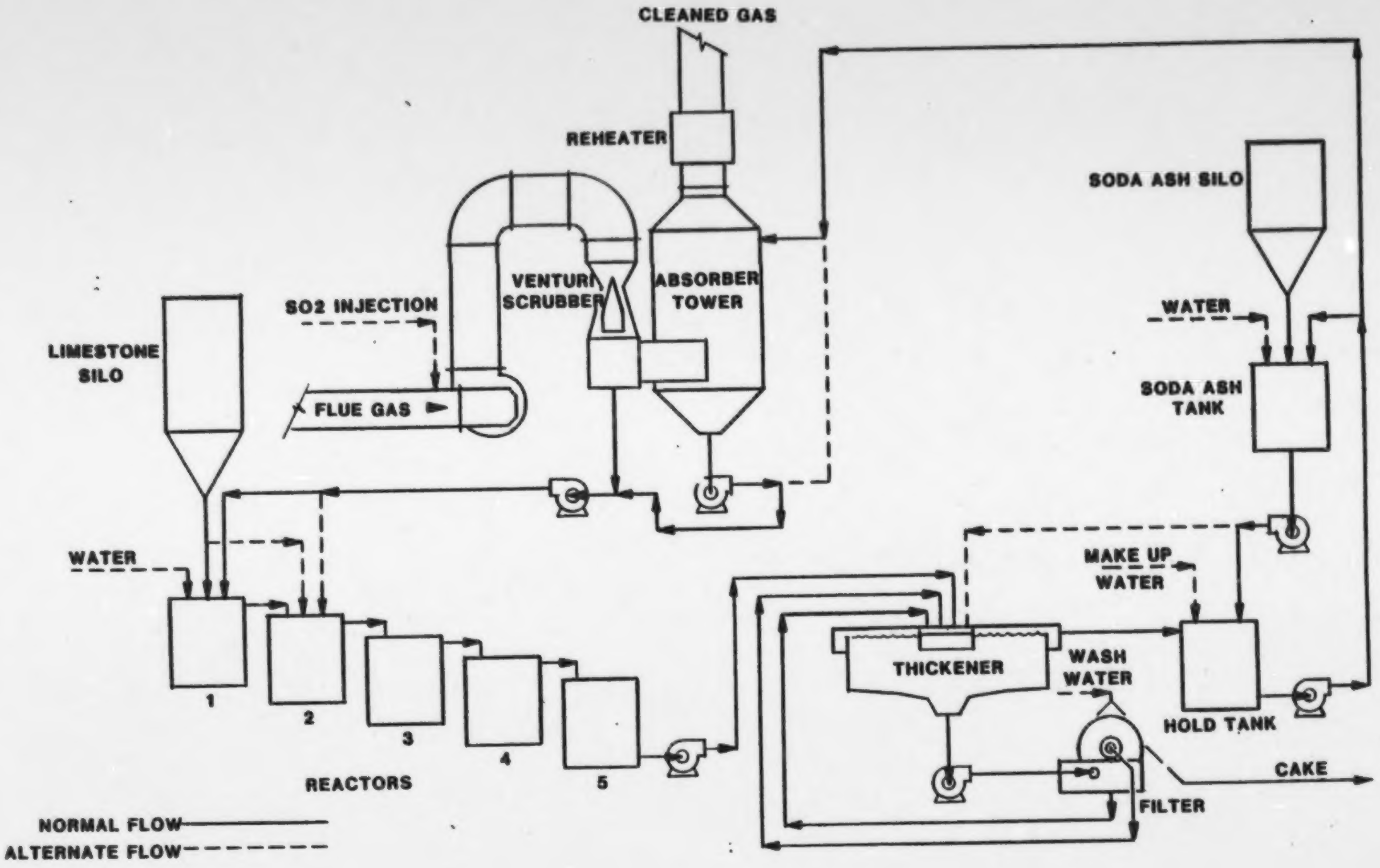

Figure 1 Plant Scholz Limestone Dual Alkali System Flow Diagram

c. Waste Solids Dewatering--

An existing 93,000 gallon thickener was utilized for solids settling. Slurry from the reactor system was pumped to the tank with clear liquor overflow collected in an existing thickener hold tank. This tank provided surge capacity for the regenerated return liquor feed to the scrubbing system. Slurry underflow was recirculated in a loop around the thickener to prevent solids settling and plugging of these lines. A bleed from this recirculation loop was fed to the existing rotary drum vacuum filter. Filtrate was returned to the thickener. Filter cake was washed on the drum using two wash water spray banks. Solids were discharged to a weigh belt conveyor for handling in the waste processing system.

d. Reagent Storage and Preparation--

Ground limestone (97-99.9% less than 325 mesh) was utilized as the regenerative reagent. It was received, stored and fed to the system in a dry form even though the capability existed for feeding it as a slurry. Soda ash solution was utlized to make up for sodium losses in the system. It was normally added to the thickener hold tank.

C. PROCESS PERFORMANCE

As stated earlier, the original intent of the test program was to evaluate several aspects of limestone dual alkali process performance. As the test program had to be reduced to only a two month period, a thorough and complete evaluation of the technology was not possible. However, process performance was evaluated for each of the following conditions:

- SO_2 removal capability
- Reagent consumption
- Power usage
- Waste product properties
- Reliability and ease of operation

1. SO_2 Removal

During the test period, inlet SO_2 concentrations ranged from 1460 to 3240 ppm (dry). Average removal efficiencies ranged from 93.5% in January to 96.7% in March.

Scrubbing liquor pH was the most important variable affecting SO_2 removal efficiency. At pH's in excess of 5.7, outlet SO_2 concentrations below 100 ppm were obtained. When operating at a pH of 6.0, outlet concentrations of less than 50 ppm were obtained. It was determined that operating the absorber at a bleed pH of 5.7 to 6.0 provided the best compromise between high SO_2 removal efficiencies and adequate regeneration of the scrubbing liquor in the reactors.

It was also determined that the ratio of bisulfite to total oxidizable sulfur compounds (TOS) concentrations in the liquor could be utilized as a guideline for SO_2 removal. The ratio of bisulfite to TOS during the test period typically ranged between 0.60 and 0.75, and this ratio produced efficiencies between 94% and 98%.

Other variables such as pressure drop, inlet SO_2, and active sodium concentration had a much smaller effect on performance than pH. Their effect appeared to be secondary in nature.

2. Limestone Utilization

Limestone utilization was very good during the test period. Utilization in the reactor train effluent was between 85% and 95% of available $CaCO_3$ in the raw limestone. Final system utilizations ranged from 93% to 100% averaging 97.5%. Fredonia limestone from Kentucky was used during the start-up and initial break-in period and Sylacauga limestone from Alabama was used during the end of the break-in period and the entire testing period. Table 3 gives the characteristics of these limestones. An attempt was made to determine the progress of limestone utilization through the five vessel reactor train. Utilization in the first reactor ranged from 23% with a level of 104 ppm solids carryover from the thickener to 63% with 1910 ppm carryover. The test report hypothesized that this was possibly due to the solids entering the first reactor acting as crystals to facilitate an increase in the rate of precipitation of calcium sulfur salts. This, it was felt, would promote an increase in the rate at which calcium was dissolved and reacted with sodium bisulfite. It was recommended that further studies be done to verify this hypothesis. Testing indicated that regardless of the extent of reaction achieved in the first reactor, efficient limestone consumption in the remainder of the train and dewatering system would occur.

3. Waste Cake Properties

a. Settling Characteristics--

The most significant process limitation encountered at Scholz was the generation of solids with good settling characteristics. During periods of the test program, solids with excellent settling characteristics were generated. However, in other periods inconsistent and poor quality solids also occured. In the months of December and February, solids that settled out to 10% of the initial slurry volume in six to eight minutes were generated. In January poor solids were generated, which took hours to settle to the same consistancy. It was determined that a significant contributor to poor solids quality was the carryover of fine solids in the thickener overflow. These solids, which were ultimately fed to the reactor after passing through the absorber, promoted the formation of finer and more difficult to settle solids in the reactor train. Laboratory work seemed to confirm this hypothesis indicating that poor settling behavior was due in large part to a high percentage of fine needle shaped solids present during operation with high solids carryover from the thickener. It was determined that solids carryover of 1000 ppm in the thickener overflow could easily be handled by the system without detrimental effects to the settling properties. However, carryover in excess of 5000 ppm resulted in rapid deterioration of solids quality.

TABLE 3. LIMESTONE PROPERTIES

	Fredonia	Sylacauga
Mine Location	Fredonia, Kentucky	Sylacauga, Alabama
Type	High-Calcium Calcite	High-Calcium Calcite
Particle Size Distribution Wet Sieve Analysis (wt.%)		
60 mesh (250)	99.7	99.9
100 mesh (149)	99.4	99.8
325 mesh (44)	96.4	97.7
400 mesh (38)	94.4	93.9
Bulk Density (lbs/ft^3)	55-60	60-65
Chemical Analysis		
Alkalinity (wt.% as $CaCO_3$)	97.9	97.8
Ca (wt.%) as $CaCO_3$)	93.6	95.7
Mg (wt.% as $MgCO_3$)	4.2	1.3
Fe (wt.% as Fe_2O_3)	0.05	0.04

It was further determined that waste solids deteriorated in their settling properties with time. On various occasions during filter downtime, solids in the thickener began resuspending as underflow was continuously recycled to the feed well. It was determined that it was possible that agglomerated, quick settling solids had begun to dissolve and recrystalize into poor settling solids.

b. Filter Cake Characteristics--

Limitations in equipment size and poor mechanical performance of the filter contributed significantly to two major shortcomings in filtering operations. These were the inability to produce filter cake with 55% or higher insoluble solids and the inability to wash the cake to reduce sodium losses to 4% of the insoluble solids content of the cake.

Leaks in the internal piping of the filter drum caused substantial losses of vacuum. This, coupled with the inability of the thickener underflow pumps to handle slurries containing more than 15% solids and the attendant dilution requirement, resulted in the filter cake having insoluble solids contents ranging between 35% and 45%.

High sodium losses in the cake were encountered throughout the test program. The system was designed with two spray banks to wash the cake utilizing up to 40 gpm of wash water. This represented the equivalent of displacing four times the final volume of liquid in a 55% solids filter cake. Corresponding sodium losses were anticipated to be approximately 4% of the insoluble solids content of the cake (equivalent to a Na/Ca ratio of 0.08 in the final cake). However, physical limitations in the handling of the filtrate, limitations in the wash water supply, and operating requirements of the waste disposal system precluded the use of more than two displacement washes. In February the average number of displacement washes was 1.5 resulting in a high Na/Ca ratio of 0.2, and in March the average number of displacement washes was only 0.8 with a correspondingly higher Na/Ca ratio of 0.4. However, there were times when up to three displacement washes were utilized resulting in sodium losses of as low as 0.03 to 0.04.

It was concluded that the problems of low insoluble solids content cake and high sodium losses in the cake were not inherent to the limestone dual alkali technology but were rather caused by mechanical conditions of the equipment used and operating requirements specific to the test facility.

4. Soda Ash Consumption

Actual sodium losses from the system at Plant Scholz were much higher than those anticipated. This was due not only to excessive loss in the filter cake as described above but also to leaks, spills and liquor purges needed to maintain volume balances. During the test period soda ash feed amounted to 0.29 moles of Na_2CO_3/mole of SO_2 removed. As much as half of this amount was needed for cake losses with the remainder associated with other liquor losses. It was felt that sodium consumption could be held within design limits of 0.04 moles of Na_2CO_3/mole of SO_2 removed on a production system with little difficulty.

5. Power Consumption

Most of the power consumption at Scholz was associated with the system forced draft fan. The venturi required pressure drop 2-3 times that of the absorber. Thus, in a typical limestone dual alkali scrubbing system where particulate removal is performed by either an electrostatic precipitator or a fabric filter, the expected power consumption would be much lower than that encountered at Scholz. Power consumption ranged from 2.5% (0.53 MW) at flue gas flow rates equivalent to a 21 MW boiler load, to 5.3% (0.42 MW) at an equivalent load of 8 MW. If the power requirement associated with the venturi is not included, a requirement of 1 to 1-1/2% of power generated at full boiler load was anticipated.

6. Process Operability

The system at Scholz was reported to have good process operability after the first two to three days of stable operation. The major problems with process operability were encountered during the initial days following any restart of the system. This was principally due to the inconsistent settling characteristics of solids generated. In some cases the solids settling rate would level off and stable operation would be achieved. At other times settling rates would continue to deteriorate resulting in significant solids carryover in the thickener overflow which further exacerbated the problem. It was hypothesized that the restart problems were caused by redissolution of solids that had been left in the thickener at the time of the outage and subsequently recrystalized into fine crystals with poorer settling characteristics.

Once the system reached stable operaton, the process operability was considered to be very good. Variations of inlet SO_2 concentrations of as much as 500 ppm were handled easily by adjustment of the feed forward rate of regenerated solution to the absorber. Variations in boiler load were handled in the same manner. In addition, upsets including carryover of fly ash in the flue gas due to precipitator malfunction, overfeeding of limestone due to operator oversite, and occasional limestone and soda ash feed outages were handled well by the system.

Some solids deposition was encountered in various portions of the regeneration system. Some scale buildup was encountered on the walls of the reactor and in the overflow pipe connecting the first and second reactors. Scale buildup in the other reactors was minimal, and it was concluded that a semi-annual cleaning operation as part of a regular maintenance program might be adequate to control any reactor scaling.

Conclusions reached as a result of the Scholz plant testing are as follows:

- A limestone double alkali process is technically feasible.
- Additional refinement and further testing of the system was necessary to develop process information required for commercial operation.
- Laboratory or small pilot plant tests should be conducted to better understand the generation of solids with good settling characteristics.

III FMC TESTWORK

Since 1977 FMC has pursued the use of limestone for regeneration in its double alkali process. They felt that the performance claims made for the process had been validated by the actual operation of three utility double alkali systems furnished by three different suppliers. The operating benefits of the double alkali technology have been demonstrated but overall evaluated costs continued to be a concern.

In 1981 FMC determined that the growing differential in the price of lime and limestone caused many evaluators to judge lime double alkali as excessively expensive when compared with conventional limestone scrubbing systems. This caused a renewed emphasis on developmental work for a limestone regenerated double alkali process. FMC's goal for a limestone double alkali process extended beyond technical feasibility. It was felt that any system developed must realize the cost benefit of using limestone while maintaining the advantages associated with lime double alkali scrubbing, and without increasing other cost components significantly. A program was developed to test limestone double alkali, with emphasis on three general operating parameters: reaction rates, solids quality, and absorber performance. Beginning in September 1981, bench scale experimental work was initiated on limestone kinetics at FMC's Central Engineering Laboratories in San Jose, California, and at Illinois Institute of

Technology Research Institute. In early 1982 the next phase of the pilot plant work was started at their Princeton Research Center in Princeton, New Jersey. The intent of the program was to define system operating ranges that allowed high utilization of limestone and commercially feasible reaction times while maintaining good solids quality and high levels of SO_2 removal.

FMC determined that the primary technical issues concerned the overall regeneration reaction. Reaction rate characteristics and the parameters that affect it were well understood from previous work done by FMC and others. It was determined that reaction rate characteristics put constraints on the process in the sense that high utilization of limestone and commercially feasible reaction times could only be accomplished with certain combinations of conditions.

It was also found that solids quality is a function of the size and shape of individual solid particles. As was seen at the Scholz test plant, these parameters were in turn a function of the chemical and physical conditions under which the particles were formed.

FMC also noted that optimum absorption of sulfur dioxide placed certain constraints on the composition of the absorbing solution, but that these conditions are not necessariliy compatible with optimum results in the regeneration reaction and the solids quality. A plan was developed to optimize the various competing factors.

A. REACTION RATES

The rate of limestone reaction and the degree to which it proceeds to completion were found to be dependent upon several parameters. First, the dissolution of limestone creates carbonate and bicarbonate ions which place an upper limit on the pH in the regeneration reaction. It is not possible to regenerate all of the sodium bisulfite without using excessive amounts of limestone. This general constraint of limited capacity for bisulfite regeneration lead to the definition of several variables that were used for defining the feasible ranges of process chemistry.

The first of these is the regeneration fraction (shown as "Y" in Figure 2). In general it was found to be difficult to achieve limestone utilizations in excess of 90% at regeneration fractions in excess of 0.6. That is, if more than 60% of the bisulfite was regenerated limestone consumption became excessive.

Another parameter defined was the ratio of the active sodium concentration to the change in bisulfite concentration in the regeneration portion of the process (shown as "Z" in Figure 2). For a given absolute amount of regeneration, the ratio is an expression of the active sodium level in the system. A high ratio corresponds to a very concentrated system and is a practical limitation based on the total dissolved solids in the scrubbing solution. A low ratio corresponds to relatively dilute solutions which lack the buffering capacity FMC normally attempts to achieve in their lime double alkali process. It was determined that scrubber pH, regeneration pH, change in bisulfite concentration, fractional regeneration, and ratio of active sodium concentration to the change in bisulfite concentration were all interrelated such that classifying any three of these parameters fixed the other two. These constraints, plus additional ones discussed below, set the optimal operating range defined by FMC for their limestone double alkali process.

Several other factors which also affect the regeneration reaction rate and limestone utilization were identified. These factors and their effects were as follows:

1. Limestone Particle Size

It was found that, as with direct limestone scrubbing, limestone particle size is directly related to utilization. Larger particles tended to become coated with the calcium sulfite reaction product rendering a portion of the limestone unavailable for reaction. Grinding the limestone to minus 200 mesh was determined to be essential for good utilization, and the relative small incremental costs for grinding to minus 325 mesh was deemed generally cost justified.

2. Limestone Reactivity

It was found that limestone varied considerably in its microstructure with a wide variety of both amorphous and crystalline characteristics. Essentially, reactivity is a surface area phenomenon with a limestone having a high surface to mass ratio being the most reactive. Therefore, it was determined that reactivity of the limestone was not necessarily defined by the mesh size.

The principal effect of limestone reactivity was determined to be on the regeneration reaction residence time requirements.

B. SOLIDS QUALITY

FMC found that solids quality is a function of particle size and shape, which are in turn functions of the chemical and physical conditions under which the particles are generated. As a practical matter, the chemical conditions are dictated largely by the kinetic and material balance constraints. Poor solids quality were generally the result of excessive nucleation, or formation of new particles of reaction products. In this sense, the goal of a fast regeneration reaction rate is in direct conflict with a goal of crystal growth. A reasonable compromise in this regard was accomplished with the proper design strategy in the reactor area. First, multiple reactors in series were utilized with split limestone feed so that the degree of reaction was controlled. Also, since the first reaction step is the most likely point to experience excessive nucleation, some solids from the thickener underflow were recycled to the first reactor, and the scrubber bleed pH kept relatively high. The former furnished seed crystals which promote particle growth instead of nucleation and the latter minimizes the driving force which also favors crystal growth.

C. ABSORBER PERFORMANCE

FMC found that the overall system material balance put constraints on the system chemistry. First, sulfur dioxide absorption began to decline dramatically for absorber pH's below 6.0. Very high collection efficiency typically requires a pH at the top of the absorber of 6.3 or higher. In contrast to FMC's lime double alkali system, however, a low pH scrubber bleed stream was necessary in order to achieve good limestone utilization. This meant that the regenerated scrubbing solution had to be fed to the top of the absorber rather than blended with the overall recirculating scrubbing liquor as is done in their lime based process. The result of this constraint was that a spray tower or packed tower proved to be the most effective absorber for a limestone double alkali process. The packed tower was deemed the most effective device because it can operate at very low liquid to gas ratios and the regenerated solution flow rate is typically equal to the total absorber flow rate requirement thus eliminating all recirculation requirements to the absorber. It was determined that a spray tower could also be utilized with regenerated solution used directly in the top stage with lower stage(s) supplied with recirculated liquor from an absorber sump.

The other important material balance constraint was found to be the practical limitation on the absolute regeneration level. The total flow rate required to regeneration is inversely proportional to the changes in bisulfite concentration achieved. For example, if the bisulfite change is 0.1 M, the flow rate required to regeneration is twice that required for a typical lime based system using a bisulfite concentration change of 0.2 M. This means that the limestone double alkali process must operate at higher active sodium concentrations in order to have regeneration flow rates equal to lime double alkali.

D. FMC PROCESS DESIGN

From their test and process design work, FMC developed an operating envelope that allows simultaneous achievement of all of the desired objectives. This envelope is summarized in Figure 2. The boundaries of the operating envelope are sumarized as follows:

1. Scrubber bleed pH equal to 5.8 - 6.6. It was found that if the absorber bleed pH was too low, the reaction rate would improve; but this was offset by an increased tendency to dissolve magnesium and a requirement for two-loop absorption. If the bleed pH was too high, only limited bisulfite regeneration per pass was achieved and a higher active sodium level was required.

2. Regeneration return pH equal to 6.3 - 7.0. It was determined that regeneration return pH must be high enough to allow good SO_2 absorption and adequate bisulfite regeneration per pass but not so high as to significantly inhibit limestone utilization.

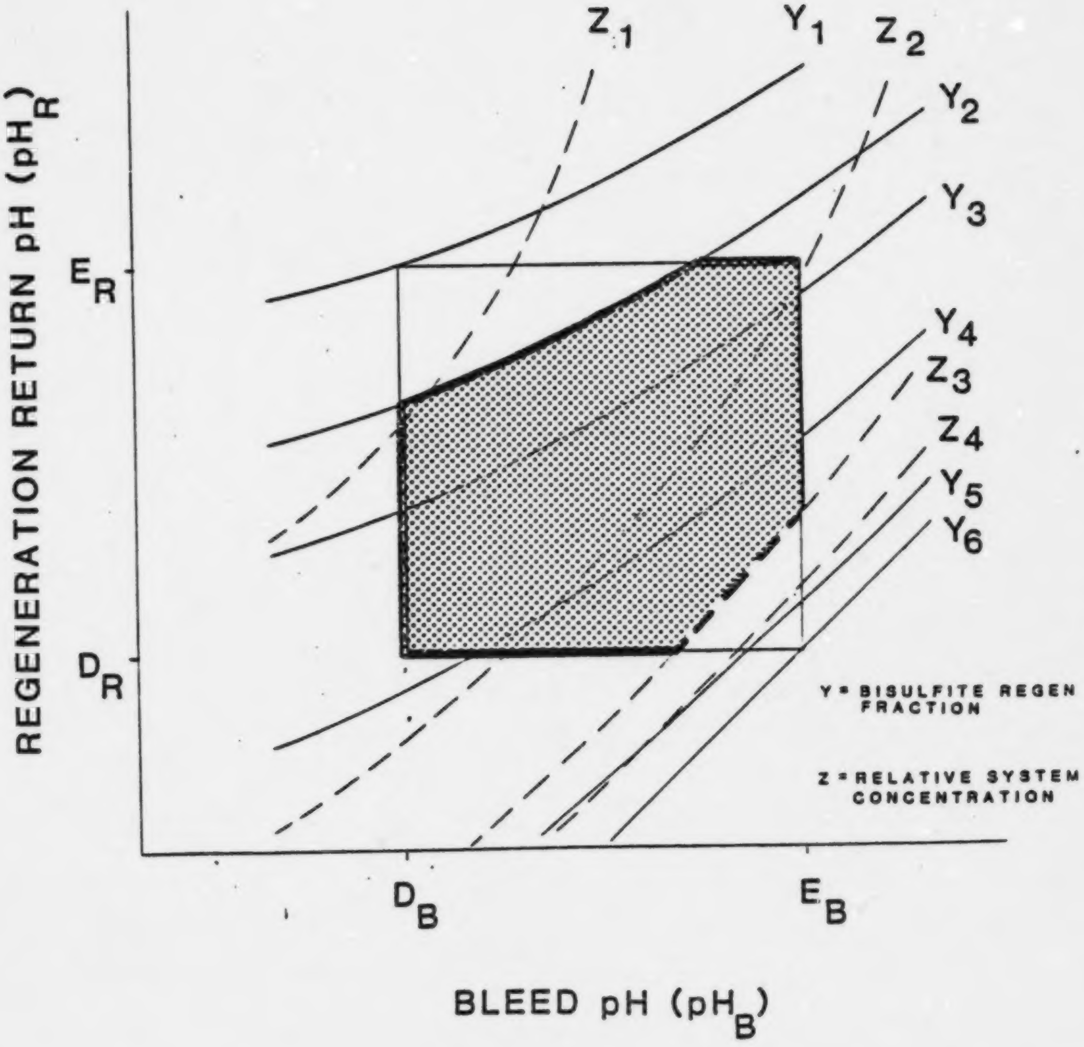

Figure 2 Limestone Double Alkali Design Envelope

3. Regeneration fraction (the percentage of the bisulfite in the scrubbing bleed solution that is regenerated in the regeneration section): FMC feels that this fraction should be in the range of 35 to 70% and preferably between 35 and 60%. Higher regeneration fractions led to poor limestone utilization, and lower regeneration fractions required excessive flow rates to regeneration.

E. DEMONSTRATION PLANT

FMC is currently in the process of installing a limestone double alkali demonstration plant at Northern Indiana Public Service Company Schahfer Station. The plant will be a 3 MWe (9,000 acfm) system installed on a slip stream from the existing Schahfer Unit 17 lime double alkali FGD system. The Schahfer Unit is designed to burn coal to 3.6% sulfur. The objective of the demonstration plant program is to confirm the limestone double alkali test results that have been obtained in the past by FMC and at the Scholz Plant.

The program schedule calls for the completion of erection by early January 1984. Testing is divided into three parts: 1) shakedown testing, 2) baseline testing, and 3) stress testing.

Shakedown testing will involve the characterization of the responsiveness of the system. The results of this period will determine those variables which most influence system operation and the appropriate values to be selected for each of these variables to attain optimal system operation. This data will be utilized as the basis for baseline testing work. The aim of the baseline test period is to demonstrate the ability of the limestone double alkali system to sustain performance over long periods of time while responding to normal changes in boiler load and flue gas conditions. Stress testing will involve establishing conditions outside the normal expected to determine the response of the system to those conditions and to determine corrective actions necessary to compensate for any adverse effects.

FMC plans to have their demonstration plant program essentially complete by the end of 1984.

IV ECONOMIC ANALYSIS

TVA recently completed an economic analysis of the limestone double alkali system for the EPA. In addition, Stearns-Roger has just finished a study for EPRI which also reviews limestone double alkali economics. The design bases and the economic procedures used in the studies differ considerably and detailed point-by-point comparisons are difficult. The major conditions and criteria used in each study are shown in Tables 15 and 16. In addition, the TVA limestone double alkali process design is based on information provided by EPA derived primarily from the tests at the Scholz steam plant described in this paper and on in-house information. The Stearns-Roger design is based largely on the process developed by FMC. In general, both studies show that the limestone double alkali process is less expensive in capital investment and operating costs than those of a forced-oxidation limestone FGD process under the conditions used in the evaluation. These results are discussed below. In addition, an ecomonic comparision of the TVA design and the FMC design, both based on TVA premises is discussed.

A. TVA/EPA STUDY*

The base case for this study utilized a new single 500 megawatt pulverized coal fired boiler firing 3.5% sulfur eastern bituminous coal. The heat rate utilized for the boiler was 9,500 Btu/kWh. The unit was assumed to be installed during the construction of the power plant and utilized a thirty year life with full load operation for 5,500 hours a year.

The system design included four 125 megawatt operating absorber trains. A spare absorber train was included to allow utilization of an emergency bypass. Hardware spares consisted of spare crushing and grinding equipment in the limestone preparation area, a spare filter, and spare process pumps in addition to the spare absorber train.

It was assumed that construction would begin in early 1981 and cover a three-year span to late 1983. Capital investment requirements were based on 1982 costs, and annual revenue requirements were based on mid-1984 costs.

In the TVA study the economics of the limestone double alkali FGD system were compared with the economics for a forced oxidation limestone scrubbing process. The major absorber process design conditions utilized in the study are shown in Table 4.

Capital Investment

The total capital investment for the limestone double alkali process was $95 million ($190/kW). The comparable forced oxidized limestone system had capital investment requirements of $103 million ($206/kW). The total direct investment, representing installed equipment costs, was $47 million for limestone double alkali and $57 million for the limestone process. A breakdown of the capital investment requirements for the two systems are shown in Table 5.

As can be seen, the largest cost differential between the systems is in the SO_2 absorption area. Here the double alkali process has a cost advantage of $9 million. While the absorbers in both processes are similar in size, the higher L/G ratio required for the limestone system resulted in a large cost penalty. The double alkali system required an L/G ratio of 3 with an L/G ratio of 106 required by the limestone process. This resulted in a recirculation pump cost comparison of $1.7 million for limestone versus $80,000 for the double alkali process. In the areas of raw material handling and feed preparation, the systems showed similar cost requirements and the regeneration area required for the limestone double alkali process did not materially add to the process cost.

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* Stephenson, C. D., Burnett, T. A., and Torstrick, R. L., Economic Evaluation of a Sodium/Limestone Double Alkali FGD Process, to be published.

TABLE 4. PROCESS DESIGN CONDITIONS

	Process	
	Sodium/Limestone Double Alkali	Limestone Scrubbing
Absorber type	Sieve tray tower	Spray Tower
Superficial Gas Velocity, ft/sec	9	10
L/G, gal/kaft3		
Presaturator/Underspray	2/1	4/10
Absorber	5	106
Stoichiometry, mole Ca/mole (SO_2 + 2HCl) Absorbed	1.0	1.4
Sulfite Oxidation, %	10	95
Thickener Feed Solids, %	1.4	8
Thickener Underflow Solids, %	25	40
Filter Cake Solids, %	55	85

The dewatering system required for limestone double alkali was evaluated at 50% higher than that required for the limestone process. This was based on the high-sulfite waste produced in the double alkali system. However, when the costs of forced oxidation were included in the limestone process, the total cost of preparing waste for disposal were similar for both systems. When a credit was applied to the limestone double alkali process for flyash disposal, the landfill disposal costs for that system were about 10% less.

The other difference in capital investment between the two processes consisted of indirect investment and other capital investment, both of which are based for the most part on direct capital investment. Exceptions to limestone double alkali's generally lower investment cost were the contingency fee and start-up and modification allowance assigned by TVA due to the lower level of technical development of the process.

Annual Revenue Requirements

Tables 6 and 7 depict the annual revenue requirements assigned to the double alkali and limestone processes respectively in the TVA study. The limestone double alkali process showed a first year annual revenue requirement of $26 million (9.28 mills/kWh) compared with $29 million (10.58 mills/kWh) for the conventional limestone process. Annual direct costs for the double alkali process are shown as $9.2 million (constituting 36% of the total first year annual revenue requirement) and for the limestone process are shown at $10.6 million (also 36% of the total annual revenue requirements).

TABLE 5. CAPITAL INVESTMENT

	Capital Investment, $	
Direct Investment	**Double Alkali**	**Limestone**
Materials Handling	2,426	2,528
Feed Preparation	4,506	4,715
Gas handling	10,800	11,281
SO_2 Absorption	11,348	20,288
Reheat	3,630	3,634
Regeneration	1,506	0
Oxidation	0	2,670
Solids Separation	5,493	3,679
Fixation	906	0
Total Process Capital	40,615	48,795
Services, Utilities and Miscellaneous	2,437	2,928
Total Direct Investment Excluding Landfill	43,052	51,723
Landfill Construction	5,247	3,781
Landfill Equipment	1,454	1,123
Landfill Credit (fly ash disposal)	(2,312)	0
Total Direct Investment	47,441	56,627
Indirect Investment		
Engineering Design and Supervision	3,014	3,621
Architect and Engineering Contractor	861	1,034
Construction Expense	6,888	8,276
Contractor Fees	2,153	2,586
Contingency	11,194	6,724
Waste Disposal Indirect Investment	1,720	1,707
Total Fixed Investment	73,271	80,575
Other Capital Investment		
Allowance For Startup and Modification	6,716	5,917
Interest During Construction	11,430	12,570
Royalties	406	0
Land	554	614
Working Capital	2,775	3,392
Total Capital Investment	95,152	103,068
Dollars per kW of Generation Capacity	190.30	206.14

Basis: 500-MW new coal-fired power unit; 3.5% in coal; 89% SO_2 removal; fixation and landfill solids disposal; north-central, new power unit with 30-yr life at 5,500 hr/yr full-load operation; 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas; reheat to 175°F as required. Mid-1982 cost basis.

TABLE 6. SODIUM/LIMESTONE DOUBLE ALKALI PROCESS ANNUAL REVENUE REQUIREMENTS

	Annual Quantity	Unit Cost, $	Total Annual Cost, k$
Direct Costs - First Year			
Raw Materials			
Limestone	107,700 tons	8.50/ton	915
Soda Ash	6,190 tons	160.00/ton	990
Lime	10,580 tons	75.00/ton	794
Total Raw Materials Cost			2,699
Conversion Costs			
Operating Labor and Supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Solids Disposal	33,280 man-hr	21.00/man-hr	699
Utilities			
Steam	528,500 klb	2.50/klb	1,321
Process Water	262,100 kgal	0.14/kgal	37
Electricity	29,618,000 kWh	0.037/kWh	1,096
Diesel Fuel	123,200 gal	1.60/gal	197
Landfill Fly Ash Credit			(312)
Maintenance			
Labor and Material			2,715
Analysis	4,990 man-hr	21.00/man-hr	105
Total Conversion Costs			6,516
Total Direct Costs			9,215
Indirect Costs - First Year			
Overheads			
Plant and Administrative (60% of conversion costs less utilities)			2,319
Total First-Year Operating and Maintenance Costs			11,534
Levelized Capital Charges (14.7% of total capital investment)			13,987
Total First-Year Annual Revenue Requirements			25,521
Levelized First-Year Operating and Maintenance Costs (1.886 First-Year O and M)			21,753
Levelized Capital Charges (14.7% of Total Capital Investment)			13,987
Levelized Annual Revenue Requirements			35,740

	M$	Mills/kWh
First-Year Annual Revenue Requirements	25.52	9.28
Levelized Annual Revenue Requirements	35.74	13.00

Basis: One-year, 5,500-hr operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE 7. LIMESTONE SCRUBBING PROCESS ANNUAL REVENUE REQUIREMENTS

	Annual Quantity	Unit Cost, $	Total Annual Cost, k$
Direct Costs - First Year			
Raw Materials			
Limestone	142,900 tons	8.50/ton	1,214
Total Raw Materials Cost			1,214
Conversion Costs			
Operating Labor and Supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Waste Disposal	29,120 man-hr	21.00/man-hr	611
Utilities			
Steam	542,200 klb	2.50/klb	1,356
Process Water	193,400 kgal	0.14/kgal	27
Electricity	57,657,000 kWh	0.037/kWh	2,133
Diesel Fuel	103,200 gal	1.60/gal	165
Maintenance			
Labor and Material			4,285
Analysis	4,990 man-hr	21.00/man-hr	105
Total Conversion Costs			9,340
Total Direct Costs			10,554
Indirect Costs - First Year			
Overheads			
Plant and Administrative (60% of conversion costs less utilities)			3,395
Total First-Year Operating and Maintenance Costs			13,949
Levelized Capital Charges (14.7% of total capital investment)			15,151
Total First-Year Annual Revenue Requirements			29,100
Levelized First-Year Operating and Maintenance Costs (1.886 first-year O and M)			23,309
Levelized Capital Charges (14.7% of total capital investment			15,151
Levelized Annual Revenue Requirements			41,460

	M$	Mills/kWh
First-Year Annual Revenue Requirements	29.10	10.58
Levelized Annual Revenue Requirements	41.46	15.08

Basis: One-year, 5,500-hr operation of system as described in capital investment table. Mid-1984 cost basis.

Levelized annual revenue requirements are \$36 million (13.00 mills/kWh) and \$41 million (15.08 mills/kWh) for the double alkali and limestone processes respectively. TVA utilized a projected value of money at 10% and an inflation rate of 6% to determine these figures.

A comparison of the first year annual revenue requirements was made and the major components are shown in Table 8. The major differences in direct costs between the systems are in the area of raw materials, electricity, and maintenance requirements. The major direct costs for the limestone double alkali process in order of significance are maintenance, raw materials, labor, steam, and electricity. For the conventional limestone process they are maintenance, electricity, labor, steam, and raw materials.

TABLE 8. COMPARISON OF ANNUAL REVENUE REQUIREMENT COMPONENTS

	Annual Revenue Requirements, k\$	
	Sodium/Limestone Double Alkali	Limestone Scrubbing
Raw Materials		
Limestone	915	1,214
Soda Ash	990	--
Lime	794	--
Total Labor	2,699	1,214
Utilities		
Steam	1,321	1,356
Electricity	1,096	2,133
Other	234	192
Maintenance	2,715	4,285
Landfill Credit	(312)	--
Direct Costs	9,215	10,554
Overheads	2,319	3,395
Operating and Maintenance Costs	11,534	13,949
Levelized Capital Charges	13,987	15,151
Total First-Year Annual Revenue Requirements	25,523	29,100

In general, the TVA study concluded that the limestone double alkali process has direct costs about 13% lower than those of a conventional limestone process primarily because of lower maintenance and power costs. Along with lower overheads which are based on direct costs--and lower capital charges for the double alkali process, the first year annual revenue requirements of the double alkali process are 12% lower than those of the limestone slurry process.

B. STEARNS-ROGER/EPRI STUDY REVIEW*

Stearns-Roger conducted a study under contract to EPRI similar to the TVA/EPA study discussed previously. The Stearns-Roger study also compared a limestone dual alkali system with a conventional limestone scrubbing system. Different design and economic criteria were utilized for the Stearns-Roger and TVA comparisons but similar conclusions were reached.

The Stearns-Roger study was based on two 500 megawatt PC fired boilers firing 4.0% sulfur coal. Table 9 outlines the specific process design criteria utilized in the cost evaluation. Estimates of raw material and utility consumptions that were utilized in the study are shown in Table 10.

Stearns-Roger also conducted a relatively subjective technical evaluation of the advantages and disadvantages of the limestone double alkali process compared with conventional limestone scrubbing. This comparison is shown in Table 11. Credit was given to the limestone double alkali process for significant advantages in maintenance as a result of reduced scaling potential, reagent reactivity, and the ability to follow changes in flue gas composition and condition.

Tables 12, 13, and 14 review the base case economic evaluation of the limestone double alkali process done in the Stearns-Roger/EPRI report. Table 12 shows the total system capital cost including the particulate removal system. Once the particulate removal system was subtracted from the total, the capital cost of the limestone double alkali FGD system was $162/kW. When compared with Stearns-Roger evaluation of conventional limestone scrubbing systems, the double alkali process requires 5% less initial investment. This is true even though a higher process contingency is included in the double alkali cost due to its lack of commercial development.

Table 14 depicts the levelized busbar cost for the limestone double alkali system. These cost are calculated by the "present worth" method (as defined in EPRI Economic Premises) and assume a thirty-year plant operating life. This levelized bus bar cost is the sum of the fixed and varible operating costs (Table 13) plus the cost of capital over the thirty-year plant life. The fixed operating costs shown for limestone double alkali are approximately 33% lower than those Stearns-Roger calculated for conventional limestone systems. This is principly due to the elimination of the high costs high maintenance of slurry recirculation pumps and the attendent lower overall maintenance requirement. Stearns-Roger also found that the total varible and capital operating costs are approximately equal for both systems.

The costs developed by Stearns-Roger in the same study for a limestone forced-oxidation process are shown in Tables 15, 16, and 17.

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* Reisdorf, J. B., Keeth, R. J., Scheck, R. W., Miranda, J. E., Economics of FGD Systems, to be published 11/83.

TABLE 9. LIMESTONE DUAL ALKALI SYSTEM
PROCESS SPECIFIC DESIGN CRITERIA*

Flue Gas Handling Area Criteria

Flue Gas Flow Rate (105% design load)	2,015,000 acfm
Pressure Drop (flange to flange)	9 in. H_2O
ESP Fly Ash Removal Efficiency	99.8%
Spray Tower Outlet Gas Temperature	127°F

SO_2 Removal Area Criteria

SO_2 Removal, High Sulfur Coal	90% (30-day rolling average)
Scrubber Design	Spray Tower, 3 Levels
Scrubber Modules	4 @ 33-1/3% capacity ea. (3 opp, 1 spare)
L/G Ratio	20 gpm/1000 acfm
Na_2CO_3 Makeup Ratio	$\frac{0.0425 \text{ mole } Na_2CO_3}{\text{mole } SO_2 \text{ removed}}$
Absorber Tower Superficial Velocity	10 fps
Reaction Mix Tank Retention Time	6 min
Oxidation Rate of Sulfite to Sulfate	18% available Na_2SO_3
Solution pH	5.8 - 6.2
Absorber Pressure Drop	6 in H_2O

Reagent Feed Area Criteria

Total Soda Ash Storage	30 days
Soda Ash 20% Solution Storage Tank	30 hrs
Regenerant Limestone Storage	60 days
Limestone Slurry Tank Storage	12 hrs
Limestone Day Bin Storage	30 hrs
Limestone Slurry Solids Concentration	25%

Absorber Regeneration Area Criteria

Limestone Reaction Tank Retention Time	120 min (total for 4 tanks)
Limestone Reagent Feed Ratio	$\frac{1.05 \text{ lb mole } CaCO_3}{\text{lb mole } SO_2 \text{ removed}}$

Waste Handling Area Criteria

Thickener Underflow Percent Solids	20%
Dewatered Sludge Percent Solids	55%
Sulfite/Sulfate Mole Ratio	9:1
Thickener Size	22 ft^2/tpd day solids
Vacuum Filter Design Basis	35 lb/hr dry solids/ft^3
Water Washes for Sodium Recovery	3

*For one 500 MW unit

TABLE 10. LIMESTONE DUAL ALKALI SYSTEM RAW MATERIAL AND UTILITY CONSUMPTION FOR TWO 500 MW UNITS*

Item	Quantity
Soda Ash @ 0.0425 Stoichiometric Ratio	2.5 t/hr
Limestone @ 1.05 Stoichiometric Ratio	61.6 t/hr
Fixative Lime @ 3% of Dry Sludge and Fly Ash	4.7 t/hr
Raw Water	1,250 gpm
Steam @ 572°F and 418 psig	221,000 lb/hr
Power (Operating Horsepower and Equivalent kW)	
Area 10 - Reagent Feed System	2,040 HP (1,520 kW)
Area 20 - SO_2 Removal System	2,695 HP (2,010 kW)
Area 30 - Flue Gas System	11,010 HP (8,215 kW)
Area 40 - Regeneration System	150 HP (110 kW)
Area 60 - Waste Handling System	2,880 HP (2,150 kW)
Area 70 - General Support Area	400 HP (310 kW)
Area 80 - Particulate Removal System	5,360 HP (4,000 kW)
Total	24,540 HP (18,315 kW)
Fly Ash (direct disposal to landfill fixated with waste sludge)	63.2 t/hr

*Operatig at 100 percent load

TABLE 11. TECHNICAL EVALUATION ADVANTAGES AND DISADVANTAGES OF THE DUAL ALKALI PROCESS WITH LIMESTONE REGENERATION*
(Compared to the Conventional Limestone Process)

Item	Strong Disadvantage	Disadvantage	Same	Advantage	Strong Advantage
1. Process					
a. Complexity of Operation			X		
b. Oxidation of Sulfite to Sulfate			X		
c. Turndown Ratio				X	
d. Load Following Capability				X	
e. Adaptability to Flue Gas Temperature Changes			X		
f. Surge Requirements			X		
g. Capability of Using Cooling Tower Blowdown			X		
h. Stability of Process			X		
i. Extreme Vessel Pressure			X		
j. High Equipment Operating Temperature			X		
k. Use of Liquid Fuel or Natural Gas			X		
l. Material Handling Characteristics			X		
m. Separation/Removal (two phase)			X		
2. Operation and Maintenance Requirements					
a. Labor Requirements				X	
b. Equipment Pluggage, Scaling				X	
c. Equipment Corrosion				X	
d. Equipment Erosion				X	
e. Reagent Reactivity/Makeup Rate				X	
3. Effect on Net Plant Heat Rate					
a. Power Consumption - Low Absorption Tower Pressure Drop				X	
b. Steam Usage			X		
4. Disposal					
a. Land Requirements				X	
b. Reactivity of Waste			X		
c. Waste Handling Characteristics			X		
5. Use of Exotic Materials of Construction			X		
6. Operational Hazard					
a. High Temperature			X		
b. High Pressure			X		
c. Use of Hazardous Chemicals			X		

TABLE 12. DUAL ALKALI (LIMESTONE) SYSTEM
TOTAL CAPITAL REQUIREMENT*

Area	Description	$/kW
20	SO_2 Removal System	41.0
40	Regeneration System	4.0
30	Flue Gas Handling System	24.0
10	Reagent Feed System	16.0
60	Waste Handling System	22.0
80	Particulate Removal System	37.0
70	General	3.0
	Total Process Capital	147.0
	General Facilities	15.0
	Engineering and Home Office Fees	18.0
	Project Contingency	30.0
	Process Contingency	8.0
	Total Plant Cost	218.0
	Allowance for Funds During Construction (AFDC)	8.0
	Total Plant Investment	226.0
	Royalty Allowance	0.7
	Preproduction Costs	7.2
	Inventory Capital	1.4
	Initial Catalyst and Chemicals	0.0
	Total Capital Requirement	235.0
	Less Capital for Particulate Control System	73.0
	Total Capital Requirement for FGD System	162.0

*December 1982 dollars, two-500 MW units.

TABLE 13. DUAL ALKALI (LIMESTONE) SYSTEM OPERATING COSTS

	1st Year $/kW*	Levelized mills/kWh
Fixed Operating Costs		
Operating Labor	1.3	0.5
Maintenance Labor	1.9	0.8
Maintenance Material	2.8	1.1
Administration and Support Labor	0.9	0.4
Total Fixed Operating Costs	6.9	2.8
Variable Operating Costs		
Lime	1.6	0.7
Limestone	5.0	2.0
Soda Ash	2.0	0.8
Raw Water	0.3	0.1
High Pressure Steam	4.8	2.0
Power	5.7	2.5
Sludge (unlined)	5.5	2.2
Total Variable Operating Costs	24.9	10.3
Credits for By-products	0.0	0.0
Total Variable Operating Costs with By-product Credit	24.9	10.3

*December 1982 dollars

TABLE 14. DUAL ALKALI (LIMESTONE) SYSTEM LEVELIZED BUSBAR COST*

	Mills/kWh
Process Capital	3.9
General Facilities	0.4
Engineering and Home Office Fees	0.5
Project Contingency	0.8
Process Contingency	0.2
Total Plant Cost	5.8
AFDC	0.2
Total Plant Investment	6.0
Royalty Allowance	0.02
Preproduction Costs	0.19
Inventory Capital	0.04
Total Capital Requirement	6.3
Fixed Operating Cost	2.8
Variable Operating Cost	10.3
Total Levelized Busbar Cost	19.4
Less By-product Credit	0.0
Total Levelized Busbar Costs with By-product Credit	19.4
Reference Case Levelized Busbar Cost	3.8
Levelized Busbar Cost Apportioned to FGD	15.6

*Two-500 MW units.

TABLE 15. FORCED OXIDATION SYSTEM TOTAL CAPITAL REQUIREMENT*

Area	Description	$/kW
20	SO_2 Removal System	67.0
30	Flue Gas Handling System	24.0
10	Reagent Feed System	14.0
60	Waste Handling System	10.0
80	Particulate Removal System	37.0
70	General	7.0
	Total Process Capital	159.0
General Facilities		16.0
Engineering and Home Office Fees		20.0
Project Contingency		34.0
Process Contingency		3.0
	Total Plant Cost	232.0
Allowance for Funds During Construction (AFDC)		8.0
	Total Plant Investment	240.0
Royalty Allowance		0.8
Preproduction Costs		7.4
Inventory Capital		0.9
Initial Catalyst and Chemicals		0.0
	Total Capital Requirement	250.0
	Less Capital for Particulate Removal System	73.0
	Total Capital Requirement for FGD System	177.0

*December 1982 dollars, two-500 MW units.

TABLE 16. DUAL ALKALI (LIMESTONE) SYSTEM OPERATING COSTS

	1st Year $/kW	Levelized mills/kWh
Fixed Operating Costs		
Operating Labor	1.0	0.4
Maintenance Labor	3.0	1.2
Maintenance Material	4.5	1.8
Administration and Support Labor	1.2	0.5
Total Fixed Operating Costs	9.7	3.9
Variable Operating Costs		
Limestone	5.4	2.2
High Pressure Steam	4.8	2.1
Power	7.8	3.3
Dry Solids (unlined)	1.7	0.7
Gypsum	1.8	0.7
Total Variable Operating Costs	21.5	9.0
Credits for By-products	0.0	0.0
Total Variable Operating Costs with By-product Credit	21.5	9.0

*December 1982 dollars.

TABLE 17. FORCED OXIDATION SYSTEM LEVELIZED BUSBAR COST*

	Mills/kWh
Process Capital	4.3
General Facilities	0.4
Engineering and Home Office Fees	0.6
Project Contingency	0.9
Process Contingency	0.1
Total Plant Cost	6.3
AFDC	0.2
Total Plant Investment	6.5
Royalty Allowance	0.02
Preproduction Costs	0.20
Inventory Capital	0.02
Total Capital Requirement	6.7
Fixed Operating Cost	3.9
Variable Operating Cost	9.0
Total Levelized Busbar Cost	19.6
Less By-product Credit	0.0
Total Levelized Busbar Costs with By-product Credit	19.6
Reference Case Levelized Busbar Cost	3.8
Levelized Busbar Cost Apportioned to FGD	15.8

*Two-500 MW units.

Stearns-Roger concluded from their study that the levelized bus bar cost apportioned to the limestone double alkali process is approximately 9% less than that of a conventional limestone process.

Discussion of the Economic Evaluations

As the foregoing summary of the TVA and Stearns-Roger economic evaluations shows, the evaluations are similar in general approach and procedure but differ appreciably in both the design and economic assumptions and specific methodology. The differences occur in three categories: the power plant size and operating conditions, the FGD system design and operating conditions, and the economic criteria and costing procedures. These differences preclude comparisons of the two evaluations in terms of absolute values or specific details. Both evaluations, however, produce the same general conclusions and illustrate the same economic relationships of the limestone double-alkali process and the limestone forced-oxidation process. This is best shown by the direct capital investment and operating cost summaries shown in the preceding discussion, selected values of which are summarized in Table 18. The absolute values of the TVA and Stearns-Roger results are not comparable because of the differences discussed, but the differences between the costs for the two processes in each evaluation can be compared.

TABLE 18. MAJOR DIRECT COSTS FROM THE TVA AND STEARNS-ROGER ECONOMIC EVALUATIONS

	TVA Evaluation		Stearns-Roger Evaluation	
	Limestone Double Alkali	Limestone Forced Oxidation	Limestone Double Alkali	Limestone Forced Oxidation
Capital Investment	$/kW		$/kW	
SO_2 Removal	23	46	41	67
Flue Gas Handling	29	30	24	24
Reagent Feed	14	14	16	14
Waste Handling	22	17	22	10
Regeneration	3	0	4	0
General	5	6	3	7
Operating Costs	Levelized, mills/kWh		Levelized, mills/kWh	
Labor	0.9	0.9	0.5	0.4
Maintenance	1.8	2.9	1.9	3.0
Limestone	0.6	0.8	2.0	2.2
Soda Ash	0.7	0	0.8	0
Lime	0.5	0	0.7	0
Steam	0.9	0.9	2.0	2.1
Electricity	0.7	1.4	2.5	3.3

In direct capital investment, the limestone double alkali process has a much lower cost for SO_2 absorption in both evaluations, while the additional costs for absorbent regeneration are relatively small. The difference in absorber costs is almost wholly responsible for the lower capital investment of the limestone double alkali process. (The Stearns-Roger evaluation also has a lower waste disposal cost for the limestone forced-oxidation process that reduces the cost difference beween the two processes; this, however, is due to the use of stacking of gypsum disposal, while the TVA design has a conventional landfill.)

In operating costs, both evaluations also shown that the maintenance costs and electricity costs are appreciably lower for the limestone double alkali process, both the result of the lower absorber costs and pumping requirements. These lower costs more than offset the additional costs for soda ash and lime.

Since the size of the absorbers is largely determined by the flue gas volume which is the same for both applications, the lower SO_2 absorption costs for the limestone double alkali process are largely a result of the smaller and less complicated absorbent recirculating system and lower pumping costs associated with the lower L/G ratio. The double alkali process is inherently flexible in the L/G ratio because the highly reactive absorbent solution--which also has a low propensity for scale formation--allows a wide choice of absorber designs. It is possible that the absorber design could range from a spray tower to a design similar to a packed tower, with corresponding high to low L/G ratios, thus allowing an optimization of absorber design and L/G ratio.

The different absorber design philosophy is seen in the designs used in the TVA and Stearns-Roger economic evaluations. The TVA design is based on information provided by EPA, in part from the limestone double alkali test at the Scholz steam plant. A sieve tray absorber is used and the L/G ratio is 8 gal/acf, 5 gal/acf of which is regenerated absorbent from the regeneration system. The Stearns-Roger design is based on information from FMC Corporation. A spray tower absorber is used and the L/G ratio is 20 gal/acf, all of which is unregenerated absorbent from the absorber hold tank.

The effects of these differences are impossible to quantify by comparing the two economic evaluations because of the differences in the design and economic premises. To compare the economics of the two designs, TVA incorporated the FMC design and operating conditions (which includes a higher limestone stoichiometry, a longer regeneration system hold time, and a lower oxidation rate, as well as, spray tower absorbers and a higher L/G ratio) into the TVA process and repeated the economic evaluation. All other conditions and economic procedures and criteria remained the same. The major conditions are shown in Table 19; other conditions do not differ from those summarized in the preceding discussion. The capital investments and annual revenue requirements are shown in Tables 20 and 21. The economics of the limestone forced-oxidation process are included for comparison.

The overall effect of the changes is small; the total capital investment is increased 4% and the annual revenue requirements are increased 4%, as compared to the original design. The use of a spray tower absorber with a higher recirculation rate increased the capital investment for SO_2 absorption by 14%, but electricity costs were decreased about 3% because of the lower flue gas pressure drop in the system. The only other capital investment components affected to a significant extent were the regeneration area costs, which increased 13% because of the longer hold time, and the fixation costs, which increased 10% because of the higher stoichiometry. The only annual revenue requirement components affected to a significant extent in addition to the electricity costs were the limestone costs, which increased 4% and the maintenance costs, which increased slightly because of the higher absorber costs.

Based on this comparison, the effects of the different design criteria on the economics of the process are minimal. The use of an absorber such as a sieve tray that allows a lower L/G ratio reduces the capital cost of the SO_2 absorption area to some degree but it has little effect on electricity costs because of the higher fan costs associated with the higher flue gas pressure drop. The changes in limestone stoichiometry, regeneration area hold time, and the oxidation rate have little effect on the economics of the process.

CONCLUSIONS

As a result of the work done by TESI at Plant Scholz and by FMC, it can be concluded that a limestone double alkali process is technically feasible. The economic evaluations by TVA and Stearns-Roger indicate that the cost of owning and operating a limestone double alkali system is less than those of conventional limestone scrubbing. On this basis, it appears that consideration of the technology is warranted by those faced with the installation of flue gas desulfurization systems, especially on plants burning relatively high sulfur fuels.

TABLE 19. MAJOR OPERATING CONDITIONS FOR THE ECONOMIC COMPARISON OF SIEVE TRAY AND SPRAY TOWER ABSORBERS

	Limestone Double Alkali	Limestone Double Alkali	Forced Oxidation Limestone
Absorber Type	Sieve Tray	Spray Tower	Spray Tower
System Pressure Drop, in. H_2O	10	8	8
Superficial Velocity, ft/sec	9	10	10
Unregenerated L/G, gal/acf	3	20	110
Regenerated L/G, gal/acf	5	0	0
Lb Na_2CO_3/lb SO_2 absorbed	0.01	0.01	0
Mole $CaCO_3$/mole SO_2 absorbed	1.01	1.05	1.4
Oxidation, %	10	5	95
Regeneration Hold Time, Min	100	120	--

TABLE 20. CAPITAL INVESTMENT COMPARISON FOR SIEVE AND SPRAY TOWER ABSORBERS

	Capital Investment, k$		
	Limestone Double Alkali Process		Forced-Oxidation Limestone Process
	Sieve Tray	Spray Tower	
Direct Investment			
Materials Handling	2,426	2,458	2,528
Feed Preparation	4,506	4,565	4,715
Gas Handling	10,800	10,615	11,281
SO_2 Absorption	11,348	12,979	20,288
Reheat	3,630	3,630	3,634
Regeneration	1,506	1,700	0
Oxidation	0	0	2,670
Solids Separation	5,493	5,533	3,679
Fixation	906	1,000	0
Total Process Capital	40,615	42,480	48,795
Services, Utilities, and Miscellaneous	2,437	2,548	2,928
Total Direct Investment Excluding Landfill	43,052	45,028	51,723
Landfill Construction	5,247	5,247	3,781
Landfill Equipment	1,454	1,454	1,123
Landfill Credit (Fly Ash Disposal)	(2,312)	(2,312)	0
Total Direct Investment	47,441	49,417	56,627
Indirect Investment			
Total Fixed Investment	73,271	76,352	80,575
Other Capital Investment	21,881	22,747	22,493
Total Capital Investment	95,152	99,099	103,068
Dollars per kW of generation capacity	190	198	206

TABLE 21. ANNUAL REVENUE REQUIREMENT COMPARISON FOR SIEVE AND SPRAY TOWER ABSORBERS

	Annual Revenue Requirements		
	Limestone Double Alkali		Limestone Scrubbing
	Sieve Tray	Spray Tower	
Raw Materials			
Limestone	915	951	1,214
Soda Ash	990	990	--
Lime	794	794	--
	2,699	2,735	1,214
Total Labor	1,357	1,357	1,269
Utilities			
Steam	1,321	1,321	1,356
Electricity	1,096	1,067	2,133
Other	234	236	192
Maintenance	2,715	2,834	4,285
Landfill Credit	(312)	(314)	--
Analysis	105	105	105
Direct Costs	9,215	9,341	10,554
Overheads	2,319	2,391	3,395
Operating and Maintenance Costs	11,534	11,732	13,949
Levelized Capital Charges	13,987	14,578	15,151
Total First-Year Annual Revenue Requirements	25,521	26,310	29,100
Levelized Operating and Maintenance Costs	21,753	22,126	26,309
Levelized Capital Charges	13,987	14,578	15,151
Levelized Annual Revenue Requirements	35,740	36,704	41,460
First-Year, mills/kWh	9.3	9.6	10.6
Levelized, mills/kWh	13.0	13.3	15.1

DEVELOPMENTS AND EXPERIENCE IN FGD MIST ELIMINATOR APPLICATION

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ABSTRACT

The purpose of this paper is to detail available mist-eliminator technology and to assess U.S. practices and experience in utilization of these sub-systems of flue gas desulfurization (FGD) processes. Pertinent trends in FGD system design and operation tied to the criticality of mist eliminator performance are identified and discussed and advancements in eliminator selection and application are reviewed. Case histories of a number of significant existing mist eliminator facilities are described and the importance of selection of internals design and arrangement permitting use of elevated mist-eliminator inlet-face gas velocity to enhance droplet separation forces is emphasized.

INTRODUCTION

The mist eliminator is a critical part of the wet flue gas desulfurization system and should be selected and effectively designed to avoid incrustation and corrosion of downstream system components, as well as to help limit the amount of carryover to the stack of suspended solids, dissolved salts and liquid (1).

FGD OPERATIONAL AND MAINTENANCE PROBLEMS RELATED TO MIST ELIMINATOR DESIGN

Results of the operation of flue gas desulfurization systems in utility plants indicate that a mist eliminator facility may be a major operational and maintenance problem, detracting from gas-cleaning system reliability. Principal difficulties have included inefficient performance, plugging/scaling, erosion and corrosion of mist eliminators, and deterioration of downstream components including stack liners.

Inadequate Eliminator Performance

Instances of poor droplet collection are the result of system design deficiencies. Under ideal service conditions, such as with provision for uniform gas flow distribution, effective means of keeping elements clean, and adequate drainoff of collected liquid, eliminators perform well. In tests conducted with mist eliminator surfaces clean and gas-flow distribution nearly uniform, extensive measurements of outlet carryover solids loadings in flue-gas discharges from 10-20 MW demonstration FGD systems with vertical-flow mist eliminators indicated that eliminator outlet solid particulate concentration was consistently lower than 0.01 lb/million Btu, and thus very much lower than the current New Source Performance Standard of 0.03 lb total solid particulate/million Btu. In all cases, scubber outlet solids concentrations were shown to be less than inlet concentrations, even for inlet solids loadings as low as 0.004 grains/SCF. Available data on commercial systems strongly suggest that poor emission-control performance by FGD mist eliminators is in some cases the direct result of fouling of the elements that changes blade profile geometry causing deterioration of efficiency.

Fouling/Corrosion

Provisions for chemical scale control designed into the scrubber slurry-recirculating system best achieve their purpose in the absorber section where dispersal of liquid flow at high rates continuously irrigates the exposed internal surfaces. Because of its lack of such irrigation, and since carryover liquid collected on and wetting the mist eliminator internal surface will absorb residual SO_2 and O_2 from the scrubbed gas, the mist eliminator is the most likely component to be fouled. Thus the surfaces of the mist eliminator must be effectively washed to prevent gypsum scaling and deposition of solids contained in sulfite/sulfate-laden droplets.

Impact on FGD Reliability

Nationwide FGD reliability data indicate (2) that the scrubbed-gas handling facilities downstream of the FGD absorbers are the principal contributors to reduced unit reliability. Critical corrosion and fouling in this low-pH wet section have been found to be strongly influenced by the extent of droplet-carryover flow at the exit of mist eliminators.

CRITICAL DESIGN ASPECTS

As noted, scrubbing systems can operate without significant deposition of solids on mist eliminator collecting surfaces. Furthermore, if the surfaces are clean and the system is properly designed, entrainment of FGD liquor does not increase outlet solid particulate emission.

Gas Flow Distribution

Uniformity of gas-flow distribution is of major significance in eliminator performance and, thus, use of design and flow-modeling techniques to equalize gas velocity across the cross-sectional flow area is essential.

One report (3) shows that local vertical gas-flow velocities entering the first level of spray banks in a spray tower ranged from approximately 120% to 20% of average velocity after the addition of integral gas-distribution vaning, as compared to a range of 250% to minus 20% without vaning. While other adjustments were called for to achieve a more satisfactory distribution to prevent poor mist-eliminator operation and liquid entrainment, it can be seen that vaned internal distribution means can be an essential first step.

In-Situ Washing Means

Although proper washing of eliminator internals is best done with fresh water, the material-balance constraints of closed-loop recirculating scrubbing-slurry systems require close limits on fresh-water input. Thus, recycled process liquid must be used in many existing scrubber systems, particularly those of the vertical gas-flow mist-eliminator type, to achieve the required mist-eliminator wash-spraying intensity. By the mechanical action of wash sprays upstream and downstream of the mist eliminator significant deposition of soft, sulfite-solids can be prevented. To control deposition of $CaSO_4$ from a super-saturated scrubber liquid the designer relies on absorber pH control and the use of $CaSO_4$-unsaturated mist-eliminator wash liquid.

Prevention of Reentrainment by Controlled Drainage

High carryover-liquid loading at the mist eliminator inlet and/or localized, high gas velocity can lead to adverse reentrainment of collected droplets when the design fails to provide adequate means of removing the liquid-catch from the eliminator elements. Through advancements in development and application of mist eliminators, improved liquid drainage provisions are available to the FGD designer.

APPLICABILITY OF ALTERNATIVE DESIGN CONFIGURATIONS

Vertical Gas Flow

Vertical-flow eliminators may be conveniently fitted into the upper section of the scrubber tower and have been preferred for this reason. They have generally used a chevron-type design with horizontal zig-zag baffling of two to six passes. Due to the tendency for liquid reentrainment the chevrons may be positioned on a slant to obtain better drainage.

Horizontal Gas Flow

The horizontal-flow mist eliminator design, most common in Japan and Germany, has only recently been applied in utility design in the United States. Baffles in a horizontal-flow eliminator are generally oriented in a vertical position. In high gas-velocity designs J-shaped phase separation chambers or disengaging columns extending from the surface of the elements counter to the gas flow afford drainage paths external to the gas-flow regime to avoid reentrainment of liquid by the gas stream. This configuration lends itself to use of higher design gas velocities with resulting increased droplet-removal efficiency due to increased inertial separating forces and better drainage. It also permits segregation of the eliminator liquid-catch from the recirculating absorber slurry stream, which is advantageous in eliminator washing-system design, permitting use of a recirculating wash water circuit.

AVAILABILITY OF GENERIC MIST ELIMINATOR TYPES FOR FGD SERVICE

INERTIAL METHODS

The most commonly applied FGD mist eliminators use gravitational (dynamic) forces caused by change in gas direction to accomplish separation of liquid droplets from scrubbed gas. A vertical-flow arrangement has been regularly used in utility service in U.S.A. Horizontal-flow design, which is more common in Japan and Germany, has only recently been introduced in utility applications here.

Chevron/Baffle Type

Vertical-flow eliminators have historically been of the rudimentary chevron, zig-zag baffling type. Per Figure 1, the sweep of the gas flow tends to reentrain collected liquid. An improvement in this design through slanting of baffles (Figure 2) provides a vertical directional component for liquid flow along the length of the baffle to give better drainage of wash water and collected mist downstream of a bulk entrainment separator of horizontal orientation designed to collect most of the inlet liquid (3). Inclining of mist eliminator baffles in this manner at 30 degrees from the horizontal (4) increases inlet liquid loading capability free of significant reentrainment by at least 100% at high vertical gas velocity. A general

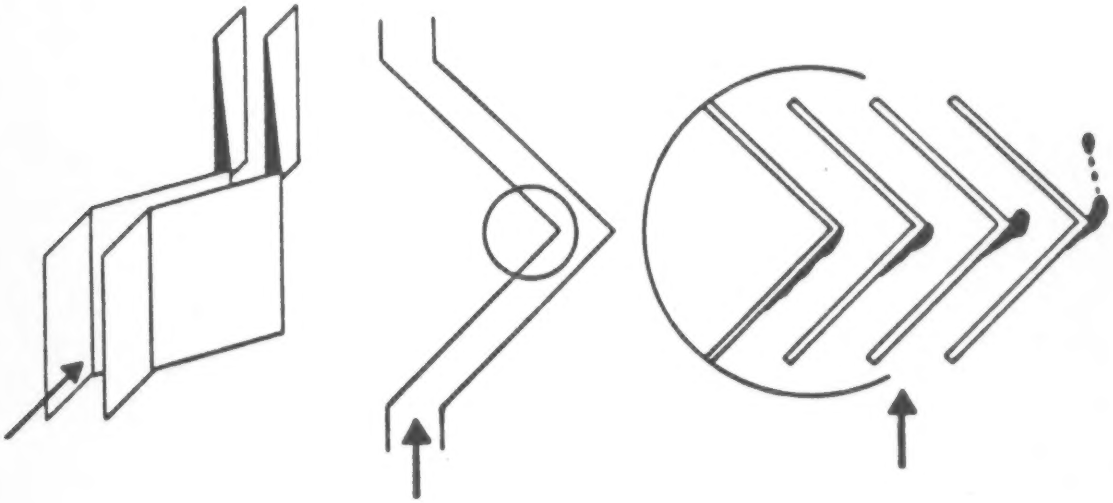

Figure 1. Reentrainment in Chevron Mist Eliminators

review (5) of commercial FGD design in the utility industry reports the following U.S. practices in eliminator application:

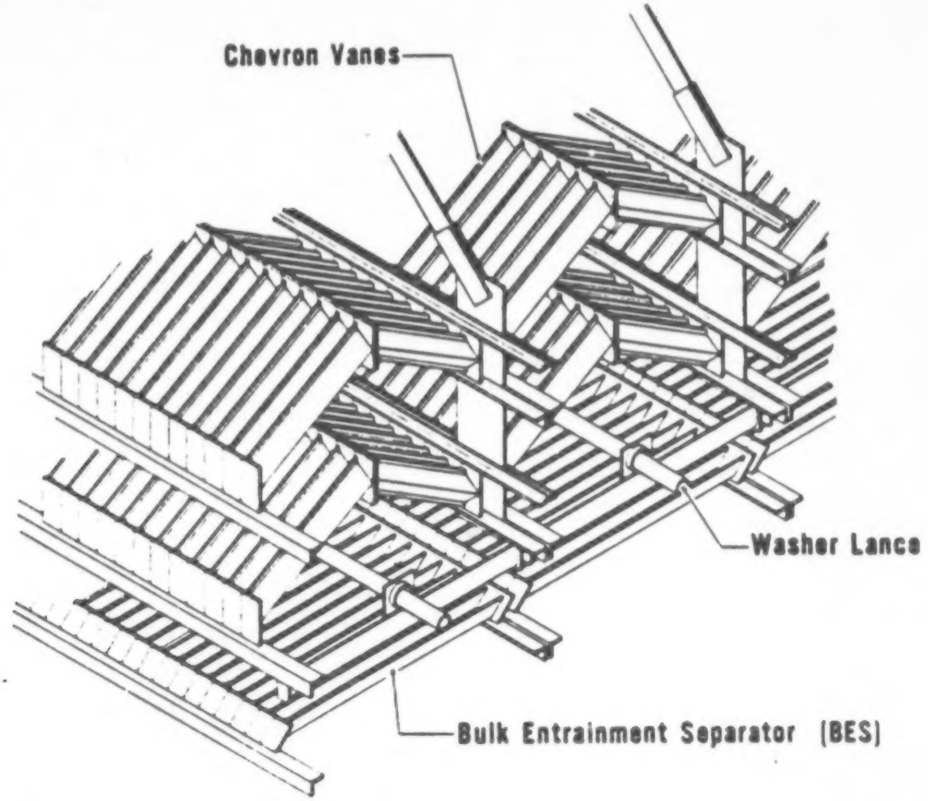

Figure 2. Slanted-Baffle Mist Eliminator

- Principal use of mist eliminators of chevron multi-pass (continuous-vane) construction in a vertical-gas flow configuration is favored from strength and cost considerations. Baffle-type units that substitute noncontinuous-slat construction are also commonly used and, like the chevron design, offer comparatively simple, open geometry with low gas-pressure drop.

- Baffle spacing is normally 1½ to 3 inches except in the second stage of two-stage designs which generally use 7/8 to 1 inch spacing.

- Plastic construction is most common due to reduced weight, cost and corrosion tendencies.

- Precollection and pre-washing stages are commonly used to improve mist eliminator operation.

- Eliminator wash systems typically operate intermittently and use a mixture of clear scrubbing liquid and fresh water makeup.

A unique and highly efficient vertical-flow baffled arrangement that utilizes exceptionally high design velocity without reentrainment is illustrated in Figure 3. This unit has advantageous 45 deg. slanting in two directions in that the total gas stream is deflected from its straight course by a series of zig-zag shaped channel walls. Slanted chevron-shaped airfoils on the eliminator walls collect and drain liquid droplets removed from the gas stream. Installations of this type operating in U.S. for over nine years have aptly field-demonstrated the high efficiency of this slant design. In some instances, a bulk entrainment separator of 45 deg

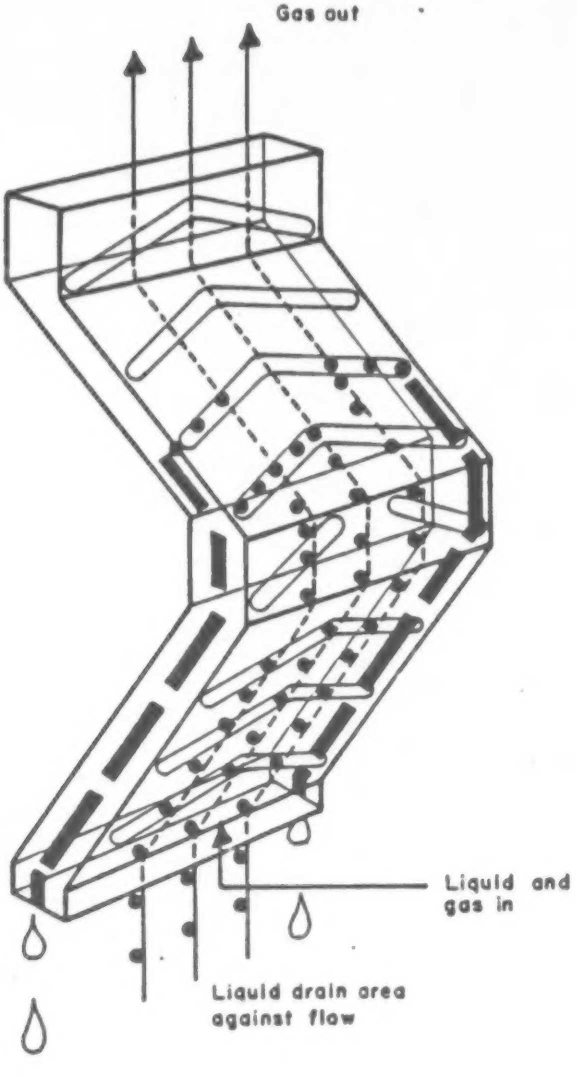

Figure 3. Mist Eliminator Drainage by Multi-Directional Slanting

orientation (Figure 4) is placed below (upstream of) the zig-zag channel-wall eliminator.

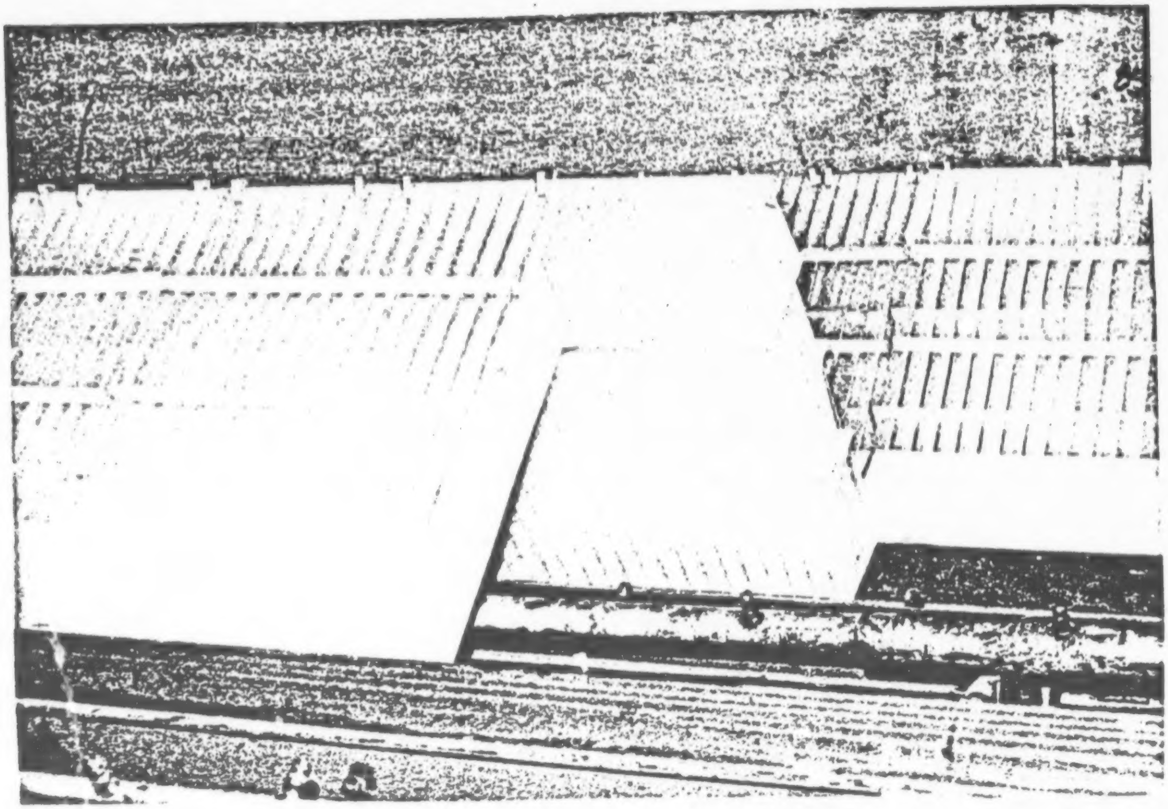

Figure 4. Slanted-Baffle Bulk Entrainment Separator

Reproduced from best available copy.

Vane Type

Elements in the horizontal flow eliminators are typically of vaned construction and oriented in a generally vertical position. As noted earlier, phase separation chambers (6) extending counter to the gas flow direction afford drainage paths external to the gas-flow regime for removal of liquid to avoid reentrainment by the high-velocity gas stream. (Figure 5).

PACKED TYPE

Neither loose nor rigid types of packing have been used for mist elimination in FGD service as they are generally prone to excessive fouling and plugging when used in services in which the flue gas or recirculating scrubbing medium have significant concentrations of suspended solids. Based on its efficient performance on small micron droplets and reliable service in difficult industrial applications, a unique layered-mesh dry-bed (7) mist

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PB

226

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84

646

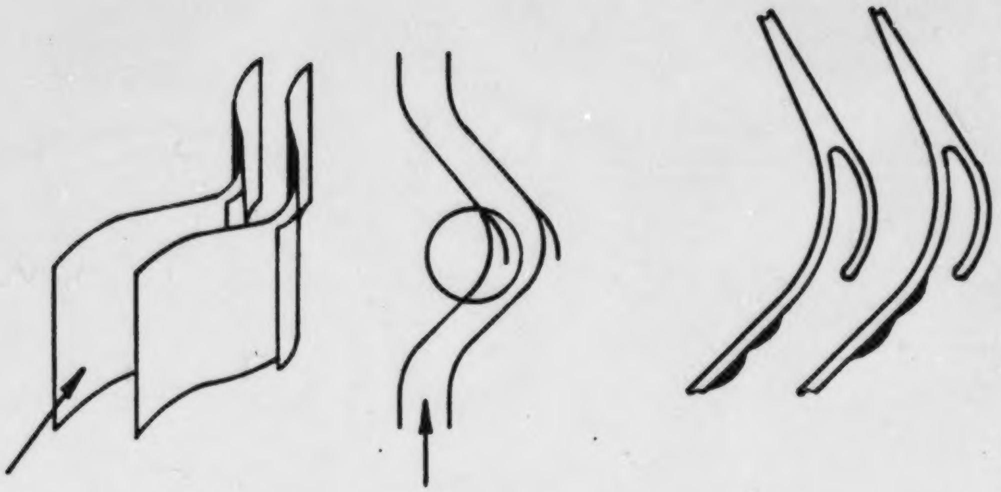

Figure 5. Function of Phase Separation Chamber at Elevated Gas Velocity

eliminator construction is now being considered for trial application in the utility industry. It is understood that the packing, an interlaced structure of thermoplastic monofilaments with essentially all of the filaments oriented perpendicular to gas flow, may generally be cleaned in-situ by intermittent spray washing. In very fouling-prone industrial services the layered-mesh beds have been found to conveniently lend themselves to periodic mechanical cleaning when temporarily removed from the unit. It is contemplated that a utility FGD scrubber module will be backfitted with a substitute triple-layer mesh-bed mist eliminator of this type to compare its reliability and performance with that of the original common horizontal-chevron type.

WET ELECTROSTATIC PRECIPITATION

Wet electrostatic precipitators, extensively used in the chemical process industries, offer a highly efficient and effective means of collected carryover droplets from FGD systems and, when present in significant concentrations, reducing emission of sub-micron sulfuric acid mist particles. Engineering and costing studies indicate (8) that the wet electrostatic type offers savings in overall stack gas cleaning system design and operation when also used simultaneously to control emissions of fine fly-ash particulate matter in fly-ash scrubbing type FGD applications that use neither dry fly-ash collectors nor venturi scrubbers. A 1 mW pilot plant test of this concept has been completed in 1983 on high-sulfur-coal utility plant service, and a 50-100 mW size demonstration is now contemplated. A unique new 150 mW cogeneration-type powerplant that is to be located at the Houston Ship Channel and fire petroleum coke has selected a wet-electrostatic-precipitator type mist eliminator for its wet FGD system in order to prevent an exceptionally high sulfuric acid mist emission tied to very high sulfur content of the fuel.

The design of the mist elimination facility is critical to FGD system dependability and to performance and economics of the utility plant itself. Available FGD mist eliminator technology has advanced broadly since earliest designs that used primitive structural angle irons and later chevrons in the scrubber outlet to reduce the liquid escape. Thus, while many U.S. system designers are content to use the simple low-velocity, low-efficiency generic eliminator configurations, advancements in inertial separation by empirically designed profiles continue to contribute significantly, worldwide, to the performance and reliability of many installations. Technologically advanced dynamic mist eliminator designs of German origin, (Figure 6 and Figure 3), utilized only sparingly in the U.S.A. to date, greatly reduce carryover that can aggravate solid particulate emissions and downstream corrosion and fouling on low-pH duct walls, flue-gas reheat devices and stack-liner walls. As noted earlier, these eliminators provide unique liquid drainage columns incorporated in aerodynamically designed eliminator vanes or passages. This permits use of elevated design superficial gas-velocity so as to more effectively trap out small micron droplets without reentraining the liquid that collects on the internal eliminator surfaces.

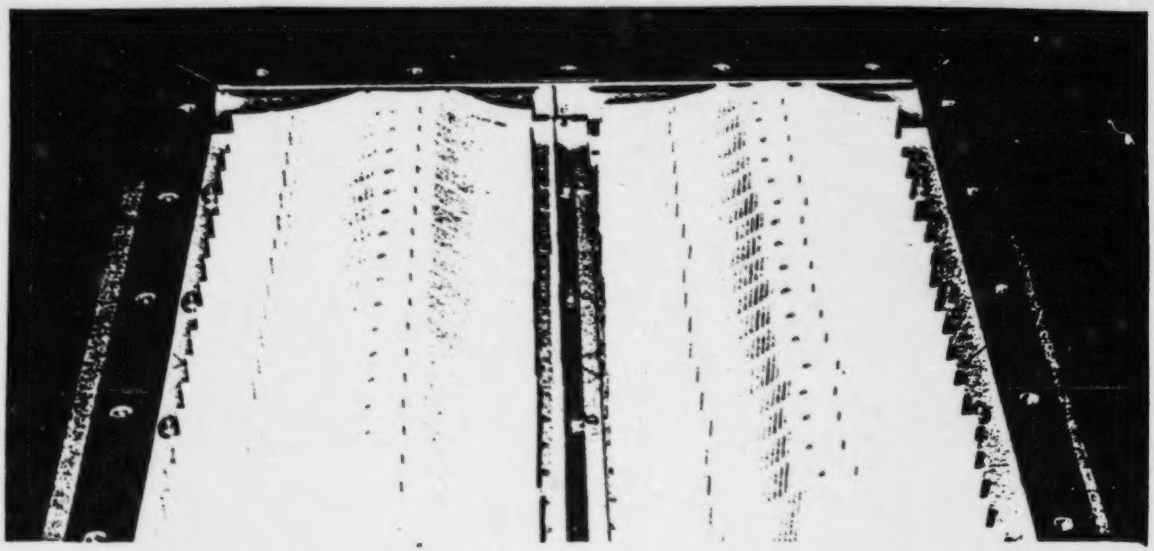

Figure 6. High-Performance Horizontal-Flow German Mist Eliminator Design

RELIABILITY EXPERIENCE AND CONCERN

FGD system reliability experience continues to be generally fair to poor and it appears, clearly, that greatest utilization of advancements in mist eliminator technology will be of significant benefit in improving the dependability of the entire wet scrubbing operation.

FGD Reliability in U.S.A.

FGD reliability, the accumulative amount of time the system was operated during any calendar period divided by the duration of time periods during which the system was called upon to operate, has averaged only 81% in the U.S.A. during the period 1974-1982 for a large number of FGD systems tracked by EPA (9). For the most recent years the average reliability has been 83.3% (1980), 82.0% (1981), and 81.3% (1982). No major improvement trend is discernable and, thus, there is substantial room and possible motivation for upgrading mist eliminator design practice and performance to yield significant benefits in increased system reliability. Moreover, in high-sulfur coal service, very low reliability, often less than 70%, has been experienced, due in part to the impact of carryover slurry contributing to outages through downstream fouling and low-pH corrosion. Selection of a high-performance mist eliminator is particularly vital in such service.

Mist Eliminator Reliability/Performance and Their Influence on FGD and Unit Reliability

Reliability of the mist eliminator depends on its being maintained in clean condition, which is directly tied to the nature of the absorber emission and the chemistry of the recirculating scrubbing medium, the quality of wash water used and the adequacy and timing of washing flows. A reliable eliminator installation that achieves high droplet removal efficiency will contribute to high FGD and unit reliability. Conversely, fouling of the mist eliminator will ultimately require shutdown of the absorber module, placing a spare module into operation if available. Additionally, FGD failure data by cause code collected from 1978 to the present by the National Electric Reliability Council (NERC) indicates (9) that FGD unreliability caused only a 2-3% forced outage rate of power-generating capacity. Thus, since as noted earlier, FGD reliability is typically only 81%, it is apparent that bypassing of FGD during FGD-shutdown, which is possible in most existing installations, has been very common. The NERC data also shows that mist eliminators were deemed directly responsible for only 9% of all plant outages caused by FGD. However, other components that are not spared, but can be significantly impacted by unremoved slurry carryover droplets, caused a substantial amount of the lost plant time. For example of all plant outages attributed to FGD failures, 28% was caused by dampers, 19% by ductwork and 17% by fans. Thus it appears that a combination of poor reliability and poor performance in many mist eliminator installations may be a major cause of FGD unreliability experienced in many FGD installations. Moreover, as discussed later, through an understanding of of the adverse chemistry and nature of emissions from U.S. supersaturated-$CaSO_4$ mode FGD installations (e.g., natural-oxidation limestone slurry scrubbing) and of critical difficulties in achieving effective washing in this mode, one can readily and clearly discern why long-term U.S. reliability experience for high-sulfur FGD installations of this mode/type has been so very inferior (10) (11) to that of unsaturated-mode installations, (i.e., magnesia-buffered calcium-based slurry scrubbing and sodium-based dual alkali scrubbing.)

Field Investigative Program by Electric Utility Industry

Due to widespread mist eliminator problems that have been reported to it, and a feeling that FGD system suppliers and architect-engineers may have been neglectful, EPRI has begun a field study program to increase understanding of how to size mist eliminators and avoid corrosion and plugging (12).

Concerns of EPRI -- Principal difficulties reported by EPRI that are to be addressed are as follows:

- Half of the existing FGD systems are probably experiencing mist elimination system problems.
- Malfunction of mist elimination systems has resulted in plugging and corrosion, causing significant FGD absorber down-time.
- The impact on mist eliminator operations of gas distribution, wash water quality and water droplet size distribution needs to be quantified.
- The effects on wet stacks of mist-eliminator operational inadequacies need to be quantified.
- There is a need for comprehensive information on mist eliminator systems so that design and operating guidelines can be established.

Scope of EPRI Work -- The EPRI project plan includes:

- Execution of field tests and data collection at specific FGD sites, including flue gas measurements, sample collection/analysis and historical data collection.
- Evaluation of mist eliminator pressure drop and velocity profile, and analysis of the source and type of plugging or carryover.
- Recommendation of improved wash sequences, water sources, maintenance procedures and design modifications.
- Development of guidelines for testing, evaluation and troubleshooting.

TRENDS IN ADVANCEMENT OF FGD SYSTEM DESIGN AND OPERATION

Broader application of high-performance eliminators is being made in conjunction with continuing advancements in FGD system design. Increasing use of spray tower type FGD absorbers has substantially raised mist eliminator liquid loadings requiring improved performance and liquid handling capacity. Droplet removal for protection against downstream fouling becomes particularly critical in systems that employ wet flue gas heat exchangers of the direct and regenerative reheat type. Current general practice in selection and specifying of mist eliminators calls for limiting droplet emission to no more than about 0.1 pounds liquid per million Btu boiler heat input, (approximately 0.05 grains/SCF, dry), thus limiting the solid emission contribution of carryover to about 0.01 pounds/MM Btu (0.005 grains/SCF, dry), which is a small fraction of the current 0.03 pounds/MM Btu New Source Performance Standard for total solid particulate emissions from coal-fired utility boilers. Sustained performance at such high efficiency is critically tied to operating conditions in the FGD system, and thus it is essential that the absorber vessel designer be cognizant of the design criteria and performance limitations of the mist eliminator. Moreover, the successful FGD installation requires closest coordination between the responsible system designer and the mist eliminator manufacturer.

Regenerative Flue Gas Reheat

The very high plant lifetime cost of energy required for adequate reheat of flue gas has led to increasing use of regenerative reheat exchangers downstream of the mist eliminator. Ljungstrom rotary heaters using hot flue gas are being employed in Japan and West Germany while tubular heat exchangers supplied with hot water are operating in U.S.A. In all such installations mist eliminator design has been a special concern, and in such applications it is particularly essential that outlet droplet loading specifications be met reliably.

Spray-Tower Type Absorbers

Because of simplified-internals and good load-turndown capability spray tower type absorbers have come into broad use in FGD service. The relatively uniform distribution of spray water across the cross-sectional gas-flow area of vertical towers served by multiple levels of sprays generally enhances gas flow distribution at the mist eliminator inlet. However, the liquid loading is generally high, perhaps higher than 1 to 2 gpm per square foot in some vertical tower designs, and efficiency potential and liquid handling capacity of alternative mist eliminator types have become significant considerations in establishing the design and staging of mist eliminator facilities.

Mist Eliminator Inlet Loadings

Inlet loadings are particularly high in limestone scrubbing systems designed for high SO_2 removal efficiencies in high sulfur service with high liquid to gas flow ratios. With high inlet liquid loading eliminator performance will generally require the inertial design to be based on use of elevated superficial gas velocity in order to achieve high-efficiency liquid removal. Advanced vertical-gas flow mist eliminator designs are adequate when meeting the outlet loading specification requires limit drop size no lower than 35 microns. Corresponding horizontal-gas-flow designs can be employed to use gas velocity sufficient for limit drop size as low as 15 to 20 microns. Moreover, generous provision is made to give adequate drainage of the liquid catch so as to avoid reentrainment.

Influence of Gas Velocity on Mist Eliminator Performance

The mist eliminator removes droplets by use of dynamically-induced inertial forces that move the liquid out of the conveying gas flow channels. Thus it becomes more efficient at increasing inlet face velocities. This is expressed quantitatively by the expression for the removal efficiency of an inertial mist eliminator:

$$E=1-e^{-z}$$

Where $z = c d_m^{x} v^{0.5}$
and where:
d_m = mass median droplet size of the inlet distribution
v = gas velocity
c and x are empirically determined values.

Typical Design -- The generic chevron arrangements that are extensively used in FGD service (Figures 1 and 2) are typically designed for nominal inlet face velocities of about 500 feet per minute yielding a removal efficiency no higher than about 98.5% at d_m = 150 microns. The high-performance vertical flow designs (Figure 7), use design gas velocities up to 1100 feet per minute, and at the same value of d_m achieve removal efficiencies exceeding 99.8%. The newest horizontal flow designs (Figure 6), use design velocities up to 1300 feet per minute, and comparable removal efficiencies are as high as 99.99%.

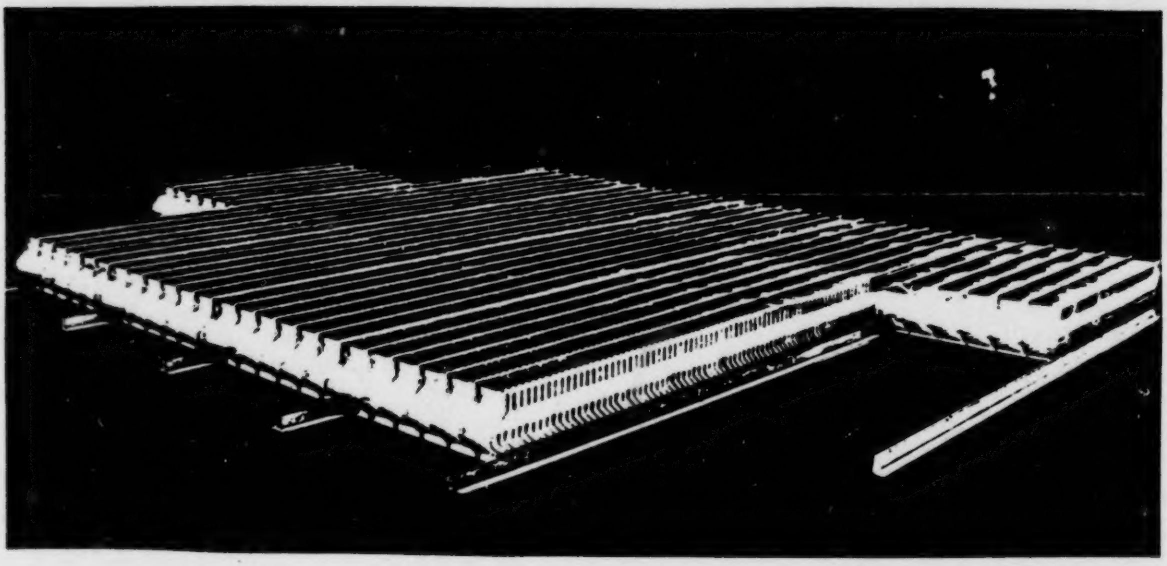

Figure 7. High-Performance Vertical-Flow German Mist Eliminator Design

Maximum Feasible Velocities -- At a given inlet liquid loading, increasing inlet face velocity requires increased capability for drainage of the liquid catch in order to avoid significant reentrainment. With chevrons reentrainment takes place at 600 to 800 feet per minute velocity. Because of high-velocity pickup of liquid (rather than drainage limitations) the high-performance, aerodynamically designed elements using drainage columns will generally encounter reentrainment above 1200 to 1400 feet per minute in vertical-gas-flow design and above 2000 feet per minute in horizontal-gas-flow design.

Gas Flow Distribution

Uniformity of gas flow distribution across the inlet face of the eliminator is critical to overall droplet collection capability. Case histories are replete with instances of poor performance coupled with mal-distribution, e.g., side-entry absorber towers with insufficient internal baffling primarily due to a lack of or inadequate gas flow modeling at the equipment design state. Inlet gas distribution is critical because at the comparatively low gas pressure-drop, e.g. generally less than 0.5 inches W.G., at which many mist eliminators operate, the eliminator is, itself, not an adequate gas-flow distribution means and cannot compensate for extreme gas-velocity gradients. Unlike other particulate collection devices, its performance is adversely affected by both very reduced gas velocity (e.g., less than 400 feet per minute) and very elevated velocity (e.g., more than 1200 feet per minute in vertical-gas-flow designs.) For optimum design to ensure that the inherent capability of the mist eliminator is utilized, the RMS (root mean square) deviation of the inlet velocity profile should ideally be in the 20% to 30% (maximum) range. Note that every zone in the mist eliminator face area at which the local velocity exceeds the level at which reentrainment starts provides a "window" through which excessive liquid discharge is emitted. Current stringent standards for total solid particulate emissions make it necessary for the conceptual design and geometry of the in-tandem absorber and mist eliminator to be established in a coordinated manner so that acceptable system performance may be realistically predicted and achieved.

Washing Technique

The key element in ensuring high mist eliminator reliability and sustained performance, and in thereby limiting impact on FGD reliability, is maintenance of day to day cleanliness of eliminator internals. Eliminator cleanliness is critically influenced by the chemical analysis of the scrubbing slurry carried over to the mist eliminator. Mist eliminator fouling with calcium sulfate scale is most effectively prevented in unsaturated-$CaSO_4$-mode scrubbing processes, i.e., those with magnesia buffering or that use sodium liquor scrubbing in a concentrated-active-alkali dual alkali system, which have dissolved calcium concentration in scrubbing liquor at the level of 170-200 ppm or less. Maintenance of an adequate level of suspended solids in scrubbing slurry and good regulation of limestone feed are important considerations in control of scrubbing slurry chemistry affecting tendencies for eliminator scaling by calcium sulfate and calcium sulfate, particularly in supersaturated-$CaSO_4$-mode limestone scrubbing systems. Of special concern in this connection are non-steady state, i.e. transient, conditions such as during load changes and at startup and shutdown that lead to elevation of scrubbing slurry pH above the set point. Use of the minimum pH in the absorber consistent with SO_2 removal requirements will help ensure the desired high limestone utilization. In-situ mist eliminator washing, when carried out effectively, provides an adequate means of removing deposits that would otherwise occur due to the presence of collected scrubbing slurry at the surface of the eliminator internals. This is best achieved when a maximum amount of FGD system makeup water is allocated to the washing activity. Constraints on process liquid discharges and on acceptable rates of FGD liquid effluent-outfall normally prevent generous use of fresh water for in-situ mist eliminator washing to ensure good surface cleanup and call for washing schemes tailored to site-specific conditions.. Based on the authors' review of available details of existing installations to be assessed in the pending EPRI investigative program noted earlier, washing effectiveness (together with gas flow distribution) may well be the key to troubleshooting success. This is particularly so in unsaturated-$CaSO_4$-mode FGD systems, which lend themselves to effective washing due to the less troublesome chemistry of the carryover droplets as noted earlier and due to the potential for satisfactorily using low-calcium $CaSO_4$-unsaturated thickener overflow liquor as part of the wash water supply. Moreover, it is clear that an assessment of the chemistry and nature of absorber emissions, unsaturated-$CaSO_4$-mode FGD vs. supersaturatedmode, performed in conjunction with an evaluation of comparative ease of mist-eliminator cleaning by in-situ washing, gives an unusually clear insight as to the principal reason(s) for poor FGD reliability experienced through 1982 in natural-oxidation supersaturated-mode high-sulfur FGD installations throughout eastern U.S.A.

Placement of In-Situ Wash Water -- Wash water is applied at the inlet face but not the final exit face of the eliminator facility. Systems incorporating two mist eliminator stages are preferentially washed on both upstream and downstream faces of the first stage and upstream, only, on the second stage.

Eliminator Washing Technique for Supersaturated-Mode Absorbers with Low Limestone Utilization -- Extensive experience has shown that absorber towers with a high concentration of limestone in the scrubbing slurry, i.e., less than 85% limestone utilization, should be provided with continuous mist eliminator wash (13). Vertical gas flow mist eliminators require 0.5 gpm per square foot, horizontal flow 0.75 gpm per square foot.

Washing When Limestone Utilization is High -- With limestone utilization 85% or greater mist eliminators should be washed intermittently, preferably with fresh or low total-dissolved-solids makeup water, 1.5 gpm per square foot for vertical flow, 3.0 gpm per square foot for horizontal flow.

Recycling vs. Once-Through Washing Systems -- In mist eliminator systems that permit segregation of used washwater and the droplet catch from scrubbing slurry, such as the horizontal-flow type, recirculated wash systems may be advantageously used. Makeup water should be comparatively fresh water, i.e. with total dissolved solids less than 1,000 ppm, and should ideally be fed at a rate sufficient to limit recirculating-water dissolved calcium level to 150 ppm and relative calcium sulfate saturation to 20%. In using once-through systems, particularly if employing recycled thickener overflow in supersaturated-mode systems, treatment with soda ash is effective in removing most of the calcium sulfate saturation in the water. The tolerable relative saturation in once-through wash liquid supply cannot be explicitly forecast as it is dependent on the type and amount of total dissolved solids.

Washing Spray Design -- Mist eliminator washing systems normally employ solid cone sprays spaced so as to provide overlapping spray patterns and designed for nozzle pressure of 30 to 50 psig. In systems where intermittent spray cycles are specified, sequence timers should provide the capability for "on" cycles of 5 to 10 minutes, "off" cycles of 20 to 60 minutes. Typically the first stage should be washed 5 minutes on, 20 to 30 minutes off, the second stage 5 minutes on, 60 minutes off. The adjustable timers should permit the cycles to be lengthened or shortened to suit operating conditions. Advisedly the wash cycle should be started at the typical rates and then be adjusted only after the system operation has been stabilized over several months. This permits use of sufficient operating experience needed to properly judge the effects of adjusting the cycle. Each adjustment should be followed by stabilized operation for several months duration to afford an adequate observation period.

Droplet Emission Testing

Measurement of must eliminator performance is of importance both for the purpose of determining liquid loadings being admitted to flue gas reheat systems or stacks and to quantify the contribution of liquid-carryover solids, dissolved and suspended, to the total solid particulate emission from the stack. In addition, in many instances the determination of the droplet size distribution is of equal importance. However, the flue gas is water saturated in the region of the absorber train in which the eliminator is positioned. Thus, significant heat losses to the ambient cause water condensation in the FGD system as well as in any gas sample trains used to withdraw and analyze gas streams. This can complicate field testing to characterize the liquid loading in gas streams entering and leaving mist eliminators and requires selected techniques to accomplish such measurements in an accurate manner.

Candidate Test Methods -- For measurement of liquid loading (14) without need to determine droplet size distribution the tester may possibly use the throtting calorimeter, heated probe, double calorimeter, condensation method or cyclone separator. When both loading and distribution measurements are required the impactor method, light scattering probes, or laser probes may possibly be used.

Criteria for Choice of Test Method -- The testing objectives that are most critical in selecting a method of carryover measurement are as follows:

- Drop size measurement discriminates among droplet diameters from 1 micron to several hundred microns.
- The method quantifies droplet concentration in the gas.
- Testing can be carried out in water saturated or unsaturated gas.
- Liquid loads in the range from a few milligrams per cubic meter to several thousand milligrams per cubic meter may be quantified.
- The measuring device must be readily portable at industrial sites.

- It must be possible in a simple way to sample and test gas across the cross-sectional flow area of large diameter vessels.

Preferred Test Methods -- All criteria, above, for selecting mist testing equipment cannot be met by any single method. Experience has demonstrated the advantageous use of two specific complementary methods/devices for measurement of size distribution and loading employing compactors -- an impactor probe of special design and a cascade impactor. The impactor probe, applicable to the size range of 15 to 2,000 microns, is a plate coated with a special film such as magnesium oxide that is inserted into the flowing gas stream via a shield-tube enclosure fashioned with a movable shutter to permit timed exposure of the plate surface. A cascade impactor, applicable to the size range of approximately 0.4 to 10 microns, comprises a group of staged in-series nozzles behind each of which is a removable thin-glass baffle (collecting) plate and through which nozzles an isokinetically sampled gas stream is drawn by a vacuum pump.

Tracer Type Testing -- In the special case of sodium liquor scrubbing systems such as dual alkali FGD a dry-filtration type collection in a filter and impingers of an isokinetically sampled gas stream (15) can be used to accurately quantify process liquor carryover loading by analysis of sodium content of the catch and relating it to the sodium ion concentration in the FGD recirculating liquor.

Materials of Construction

Mist eliminator users are increasingly emphasizing the need for capability to withstand upset or transient temperatures ranging as high as from 350° F. to 700° F. This has brought about rapid development and use of newer high-temperature plastics as well as high-alloy metals.

Materials with Modest Temperature Capability -- The lower temperature plastics include PVC (polyvinyl chloride), polypropylene and phenyele oxide based material. The latter contains amounts of polymers such as styrene added to make it more extrudable or moldable and is made in various grades identified by heat deflection temperatures as high as 300° F. and more. However, when subject to stress and heat in actual service, the maximum service temperature is considerably less than the heat deflection rating.

Materials for Elevated Temperatures -- Particularly as a result of proximity of heat sources such as flue-gas reheat means, positioned close to the mist eliminator in some applications, high temperature plastic materials such as FRP (fiberglass reinforced polyester) are increasingly being specified and used for mist eliminator internals. FRP can typically operate up to approximately 350° F., and in some cases, with some charring, has tolerated temperatures approaching 500° F. The pultruded blade (profile) forms typically contain a mixture of premium-grade isophthalic polyester resin, alumina/silicate filler and approximately 5% antimony trioxide serving as a fire retardant. The addition of a synthetic cloth surfacing veil provides protection for the glass reinforcing fibers.

OUTLOOK FOR FUTURE FGD SYSTEMS APPLICATION

The strong trend in U.S. since 1978 toward application of forced-oxidation type calcium-based FGD systems in preference to previously common natural-oxidation supersaturated-mode types may be expected to somewhat reduce absorber-emission related impacts on mist-eliminator and FGD reliability. Nonetheless the forced-oxidation processes do also, to a degree, operate in the supersaturated mode. Those being applied in U.S.A. may be expected to pose significant fouling problems for the eliminator step and low-pH corrosion and fouling immediately downstream of the mist eliminator due to many circumstances not encountered in the high-reliability unsaturated-mode systems previously discussed. These include:

- Presence of unreacted lime/limestone in carryover liquid causing fouling, particularly when the goal of high system limestone utilization is achieved by use of a precooling or other remote scrubbing step
- Comparatively high carryover rates due to the high absorber liquid/gas flow ratios required for SO_2 removal and/or gypsum scale control in the absorber causing fouling/corrosion
- Comparatively high outlet flue-gas SO_2 concentration due to inherently lower SO_2 removal efficiency causing corrosion
- Potentially higher outlet flue-gas HCl, HF, NO_2, and SO_3/H_2SO_4 concentration (due to a possibly lower absorption mass-transfer capability) causing corrosion.

CONTINUED MISAPPLICATION OF INERTIAL MIST ELIMINATOR TECHNOLOGY

Notwithstanding the critical need for improved standards of eliminator performance and reliability, current mist-eliminator specifications of architect engineers and FGD system suppliers frequently call for use of very low design gas velocity, e.g., no greater than 480 feet per minute with no exception allowed, thus sustaining the use of the simple low-efficiency chevron designs. In some cases it appears that such requirements are set by structural engineering considerations rather than the needs of the process design. Further, in some instances where somewhat higher design gas velocity is used, chevron-type equipment is employed at design gas velocity constituting 80 to 85% of the critical (reentrainment threshhold) velocity. Performance of such an installation is drastically affected by wide deviations in local gas flow rate. Thus, the lack of attention to or the disregard for optimal mist eliminator application will continue to result in operating FGD installations that fail to meet system performance and reliability objectives. Success of an FGD system is obviously dependent on close coordination between the Engineer responsible for system design and the mist eliminator applications-specialist/supplier.

IMPROVED INSTALLATIONS FOR REGENERATIVE AND DIRECT STEAM REHEAT

Control of scrubber carryover plays a direct role in all FGD systems in helping to limit fouling and corrosion of downstream components handling wet gas. Moreover, mist eliminator performance and reliability is particularly critical in achieving dependable operation of flue-gas reheat heat-exchange equipment that is in direct contact with the scrubbed gas. Use of high-efficiency mist eliminator design to minimize droplet carryover (in conjunction with employment of in-situ high flow-volume steam or air blowing of the most exposed portions of the gas-wetted heat exchange surface to periodically remove deposits) has been successful in limiting exchanger corrosion and maximizing heat transfer.

Regenerative Reheat

In current application of FGD in high-fuel-cost nations like Japan and West Germany growing use is being made of means of utilizing heat in raw flue gas exitting electrostatic precipitators for reheating of water-saturated flue gas from wet scrubbers. Ljungstrom rotary heat exchangers are applied to achieve major plant lifetime-savings by elimination of use of reheat heat-sources requiring firing of additional fuel. A tubular-exchanger type regenerative reheat system has been utilized in U.S.A. in an FGD facility now in initial startup. Like the Ljungstrom units, this first-of-a-kind facility indirectly cools the raw flue gas below its sulfuric acid dewpoint temperature in extracting "free" heat for flue gas reheat.

TVA, Paradise Steam Plant -- The regenerative-reheat-equipped 704 mW Paradise Unit No. 1 high-sulfur forced-oxidation FGD system, Drakesboro, Kentucky, has been undergoing startup and commissioning throughout 1983. A two-stage high (1500 feet per minute) velocity horizontal-flow mist eliminator operating at a 120° F. water-saturated condition specified for approximately 0.07 grains/SCF (dry) outlet liquid loading (16 micron limit drop size) is utilized immediately upstream of a Type 317L tubular stainless steel reheat exchanger designed to cool recirculating regenerative-reheating water to 188° F. A downstream tubular carbon steel second-stage reheat exchanger is designed to receive the hot water supply at only 200°F. and further reheat the flue gas to a reheater outlet temperature of 170° F. Although only approximately 50% of regenerative flue-gas reheat capacity has been realized in the 1983 operation to date, the mist eliminator operation and use of retractable steam blowers at the reheat exchanger has prevented any deterioration of the stainless steel tubes. Because of the temporarily inadequate reheating, the carbon steel reheat exchanger tubes have been attacked by sulfuric acid deposition for which the system was not designed. Due to inadvertant overheating of shut-down absorber modules by hot water allowed to flow through the reheat exchangers, meltdown of mist eliminator internals has occurred on several occasions, which may require replacement with FRP.

NWK, Wilhelmshaven Power Plant -- The Ljungstrom rotary-regenerative reheater equipped 350mW No. 2 FGD module at the low-sulfur Wilhelmshaven, West Germany, plant has been in sustained high-availability operation since early 1982. A two-stage high-velocity horizontal-flow mist eliminator operating upstream of the regenerative exchanger at a 115° F. water-saturated condition was specified for approximately 0.09 grains/SCF (dry) outlet liquid loading (20 micron limit drop size) and has been discharging only 0.03 grains/SCF (dry) by test. The system design and initial operation provided for 10 to 15° F. reheat by flue-gas by-pass upstream of the Ljungstrom, which heats the flue gas an additional 90° F. The prior reheating is done so as to help ensure against excessive sticking of mist eliminator carryover slurry in the internals of the regenerative exchanger. In a current trial run with by-pass reheat shut off, the FGD system had at last report been operating for eight weeks without any problems with regenerative reheat.

Direct Reheat

Direct reheat of flue gas is accomplished by use of an in-line tubular heat exchanger that is installed directly in the scrubbed flue-gas stream and heats the gas by indirect heat exchange with steam or hot water. The heat exchanger is particularly vulnerable at its gas inlet end where the gas-wetted surface is subject to low-pH attack. Crevice corrosion is especially troublesome if these surfaces become continuously fouled with solids originating from scrubbing slurry carryover droplets. In a number of installations improved mist eliminator designs have been successful in limiting carryover flow and, with periodic tube blowing, avoiding significant reheater corrosion.

Arizona Public Service Co., Cholla Station -- No. 2 FGD system, low-sulfur coal service, utilizes a modern-design vertical flow mist eliminator as in Figure 7 to protect a steam heated flue-gas reheat exchanger that contains Incoloy alloy 825 tubes. In conjunction with intensive periodic steam blowing of the inlet face of the reheater bundle, the mist eliminator has been fully effective in preventing fouling and corrosion of the reheater.

Montana Power Company, Colstrip Station -- Utilizing a wash tray with a very high (3" to 5" W.G.) gas-pressure drop ahead of chevron mist eliminators the low-sulfur FGD systems at Colstrip have operated without significant fouling or corrosion of the steam-heated direct-reheat exchangers, which are plate coil type constructed of Inconel alloy 625 (first stage) and Hastelloy alloy G (second stage). As in the case of Cholla Station, intensive periodic steam blowing of reheater internals are essential to sustained reheater reliability.

RETROFIT UPGRADING OF COMMON LOW-EFFICIENCY MIST ELIMINATORS

In numerous instances, and in order to upgrade performance, reliability and temperature rating the advanced vertical-gas-flow mist eliminator design (Figure 7) has been retrofitted in FRP to existing modules utilizing space originally occupied by chevron type installations.

Big Rivers Electric Cooperative, Green Station

The original chevron elements suffered from meltdown during brief shutdown periods due to proximity to the flue-gas reheat heat source. In retrofitting the advanced vertical-flow design, area-by-area replacement was required by the Engineer and, hence, no increase in superficial gas velocity nor advantageous reduction in limit drop size was accomplished. However, field observations indicated that the appearance (opacity) of the stack-discharge trailoff was improved.

Duquesne Lighting Company, Phillips Station

A similar replacement was made at Duquesne. Operation is understood to be satisfactory.

Louisville Gas and Electric Company, Cane Run Station

Retrofitting of the same advanced design together with a compatible bulk-removal mist eliminator (Figure 4) upstream of it was motivated, in part, by incidents in which prior chevron elements melted down during shutdown periods due to proximity to the reheat heat source. Through an eliminator washing improvement program a substantial increase in sustained performance and reliability has been achieved.

CONCLUSIONS

1. Mist eliminator selection and design as well as and washing technique are major factors influencing the performance and reliability of FGD systems.

2. Mist eliminator technology advancements in design and forming of aerodynamically shaped, extruded profile elements for use in either a horizontal or vertical-gas-flow regime offer means of gaining potential performance substantially greater than that of chevrons.

3. Absorber design provisions for gas flow distribution are of critical importance in realizing the performance capability of an eliminator system.

4. Satisfactory mist eliminator performance and reliability will be best achieved when the designs of the absorber system and of the mist-eliminator system are coordinated through the integrated efforts of the wet scrubber designer and mist eliminator supplier/specialist.

5. Traditional use of low-velocity chevron mist eliminators continues and is judged to be an important contributing factor in many low-pH-end FGD reliability problems and in instances of poor or marginal gas cleaning performance.

6. Conservative, high-performance mist eliminator design and effective eliminator washing operation are of critical importance in supersaturated-$CaSO_4$-mode FGD systems, both natural and forced oxidation, in view of serious impact on FGD system reliability of the absorber-emission species characteristic of this type of process.

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FGD GYPSUM: UTILIZATION VS. DISPOSAL

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ABSTRACT

The purpose of this paper is to give a technical and economic evaluation of alternative means of managing wastes collected in forced-oxidation (FO) type flue gas desulfurization (FGD) systems. Worldwide application of FO-FGD is described and environmental considerations in selection of waste management alternatives - gypsum utilization vs. disposal - are reviewed. Feasibility of and barriers to commercial gypsum from FGD in North America during the 1980s are analyzed. The impetus, method of implementation, and potential for gypsum use are addressed for major by-product gypsum FGD installations now in operation, under construction or in the design phase. It is concluded that in many instances, the production and sale of usable gypsum from FGD would be a benefit and a source of direct profit to the utility plant owner, but that in most cases the gypsum depletion allowance under IRS Code 613B will continue to be a major disincentive for purchase of by-product gypsum by vertically integrated gypsum companies in U.S.A.

INTRODUCTION

PRODUCTION OF GYPSUM BY FORCED OXIDATION

Gypsum, double-hydrated calcium sulfate, is a naturally occurring non-metallic mineral used as a raw material in the manufacture of gypsum board, Portland cement, plaster products and in agriculture. Forced oxidation as applied to wet lime/limestone flue gas desulfurization (FGD) consists of forcing air into the spent scrubbing slurry to produce gypsum, the oxidized form of calcium sulfite. The gypsum can then be easily dewatered to a cake containing greater than 80% solids by weight. This operating technique is commonly used worldwide. North America is the last major FGD-using area to adopt FO operation beginning in 1978, bringing about a major departure from the natural-oxidation design and operation of FGD facilities. Since that time, approximately 20 power plants have purchased provision for FO in FGD operations constituting 16,000 mW of scrubbed generating capacity and almost half of all limestone FGD committed to date.

ADVANTAGES OF FORCED OXIDATION

This SO_2 removal method is superior to natural-oxidation FGD commonly applied in U.S.A. since it provides improved control of FGD-process operating conditions, which may result in higher FGD availability in limestone based SO_2 absorption. It results in improved control of internal scaling, better utilization of limestone chemical reagent and enhanced SO_2 removal efficiency.

Moreover, the solid waste produced from the SO_2-catch is in the form of comparatively coarse-grained calcium sulfate solids that may be more easily dewatered, transported and stored and which require less than 60 percent of the solid waste disposal area required for direct ponding of sludge without forced oxidation. Additionally, gypsum formed by FO-FGD has potential for use in gypsum board manufacture, cement manufacture and agriculture.

MANNER OF WORLDWIDE IMPLEMENTATION

On a worldwide basis, wet scrubbing using FO of the SO_2-catch is the principal means of desulfurization of flue gas (FGD) from electric utility boilers. FGD waste solids, initially generated as a slurry, are subsequently dewatered to a sludge or cake containing both solid and liquid waste forms. The decanted liquid and filtrate is generally returned to the process. However, build-up of dissolved solids in the scrubbing slurry may preclude closed-loop operation resulting in a wastewater effluent requiring disposal. Production and use of salable gypsum avoids discarding of FGD solid waste at a permanent disposal site.

The production and utilization of a usable grade of gypsum from FO-FGD is routinely done in Japan and West Germany, the two principal countries that have applied flue gas desulfurization overseas. However, in view of the limited number of usable-gypsum type limestone-scrubbing systems that have been purchased in North America (see Table 1), we continue to be principally oriented to throwaway-solid-waste FGD operation even after the transition in the past five years to use of forced oxidation instead of natural oxidation.

DISPOSAL OF FGD GYPSUM

The disposal of FGD wastes, both dewatered solids and liquid effluents, has the potential for pollution of groundwater and surface water bodies impacted by waste disposal activities.

WASTE SOLIDS

Land-based disposal of scrubber solids from solids-precipitation-type FGD systems is a problem principally because of the very large waste quantity generated. At the same time, EPA regulatory activity is concerned with the possible long-term effects leading to pollution of surface water and groundwater bodies due to the leaching of constituents such as trace-metals

TABLE 1. COMPARISON OF COMMITTED FO-FGD INSTALLATIONS YIELDING

SALABLE GYPSUM VS. THROWAWAY SOLID WASTE

Power Plant	mW	% of Total
AESC, Cogen. 1, Houston, Texas	150 (Approx.)	1.0
JEA, SJRPP 1 and 2, East Port, Florida	1,200	7.7
MPW, Muscatine 9, Muscatine, Iowa	166	1.1
Tampa Electric, Big Bend 4, Tampa, Florida	475	3.0
Sub-Total, Salable Gypsum Producing:	2,991	12.8
Aggregate of FO-FGD Throwaway Waste Installations:	13,517	87.2
Total FO-FGD Committed in U.S.A.	16,508	100.0

from these waste solids. However, hydrogeology of typical sites that may be used for storage or ultimate disposal of wastes from such high-volume sources can be complex. Thus, for an individual site there may be uncertainties as to the extent of adverse environmental impact due to permeation of leachate through the discarded waste mass and the underlying strata. The two main disposal options for discarding solid waste generated by FO-FGD are landfilling and wet stacking.

Landfilling

FO-FGD operation yields typical gypsum crystal size of from 40 to 100 microns, producing an easily dewatered and handled coarse-grained cake that can, if necessary, be readily transported as an 80-90% solids mass to a remote disposal site. However, an implication of this low moisture content, (and the inability of the discarded coarse-grained waste to retain any significant amount of FGD liquor over an extended period), is that little of the scrubber liquor can be disposed of by occlusion in the solid waste mass. Consequently, the operation of closed loop FO-FGD may not be possible because of an increase in system corrosion resulting from a greater than 10-fold increase in FGDliquor dissolved solids concentration e.g., to in excess of 100,000 ppm chloride in the case of a medium/high sulfur coal application (1).

Environmental concerns associated with landfilling of FO-FGD waste solids include:

- Water-soluble material (dissolved metals and salts including calcium sulfate) is present in seepage and runoff.

- After the lifetime of the FGD operation, seepage and runoff of leachate will continue unabated with no means of assimilating it by recycle operation. (The substantial solubility-in-water of the gypsum solids will yield an endless source of leachate saturated with calcium sulfate and containing other components such as sulfite, dissolved metals and salts.)

Wet Stacking

Based on field tests (2), FO-FGD solid waste may instead be disposed of by wet stacking. This is a technique utilized extensively in the phosphoric acid industry in Florida for land-based disposal of waste-gypsum, wherein a 10 to 15% solids aqueous-slurry is discharged at a disposal site so as to permit formation of a sharply-pitched stack over the plant lifetime. In this method to be applied at a number of U.S. power plants, reduced acreage use is achieved through such stacked disposal without need for dewatering by thickeners and filters or centrifuges. Overall capital costs and annual revenue requirements are understood to compare favorable with even simple ponded-waste natural oxidation FGD systems. In the so-called "upstream method" of construction, the wet stacking is begun by first constructing an earthen starter-dike forming a sedimentation pond. FO-FGD solid waste is pumped to the pond in a slurry form, usually at 10 to 20 percent solids, and allowed to settle and drain (3). Process water is decanted from the pond and returned to the plant. When sufficient gypsum is deposited within the pond, gypsum is excavated with a dragline to form and step-wise build up a perimeter dike immediately inside the starter dike. Decanted liquor is piped under the starter dike to an annular perimeter ditch and surge pond formed by the starter dike and a clay exterior dike. The process of sedimentation, excavation, and raising of the perimeter dikes continues on a regular basis during the active life of the stack. Using the upstream method of construction, gypsum stacks in the phosphate industry have reached heights exceeding 100 feet (30.5 m) with slopes of 1.5 horizontal to 1.0 vertical, which is approximately the angle of repose of some gypsums. These steep slopes result from casting the gypsum with a dragline and allowing some gypsum to roll down the outside of the stack to eliminate need for shaping. Therefore, the gypsum perimeter dikes of some stacks have a factor of safety very close to unity and from a conventional geotechnical engineering point of view, failures of gypsum stacks sometimes occur. Fortunately, gypsum is a very forgiving material, and, unlike most mine tailings, gypsum does not readily flow. Therefore, the consequence of a failure is the loss of process water stored on the gypsum stack. If process water escapes the plant property, liabilities from pollution and environmental damage may also result. On a day-in, day-out basis, seepage of process liquid to groundwater and surface water should be controlled by use of an impervious foundation as well as of clay-type exterior and starter dikes, and by collecting and treating or recycling all seepage and runoff. Although a greater occlusion of process liquor in solid waste occurs in wet stacking than in landfilling, in situ washing of the stacked solids will occur (due to the high permeability of the stack). This leachate when subsequently collected and recycled would be expected to lead to a salt concentration build-up even greater than that which would occur using in-plant dewatering and landfill disposal. This would most certainly preclude closed loop operation of the FGD system in the case of wet stacking.

Environmental concerns in use of wet stacking for a FO-FGD waste include:

- The very large working area of the stack facility will cause the liquor recycle to FGD (returning from the stack-pond section) to also return large quantities of liquid running off from the entire stacking facility. Resulting water-balance problems in rainy weather may lead to intermittent outboarding of substantial discharges of process-liquor.

- All pile run-off will be saturated with calcium sulfate and contain gypsum surface-moisture constituents (dissolved metals and salts).

- Over a twenty-year lifetime, a large stack in excess of 100 feet high may be created, e.g., an increment of 40 mW, 3 percent sulfur bituminous coal, will generate a simple truncated conical stack, 400 feet in base diameter, 100 feet in diameter at the top, 100 feet high, with side slopes of 1.5 horizontal to 1.0 vertical. In no country are regulations or experience available as to how to satisfactorily decommission a lifetime gypsum stack or how to reclaim it.

- Stack slopes are too great to hold topsoil and vegetation. Thus, normal concepts for "covering" of the waste in this manner do not apply.

- After the FGD lifetime, seepage of leachate will continue unabated with no user to which to recycle for re-use. The substantial solubility-in-water of the gypsum solids will yield an endless source of leachate saturated with calcium sulfate plus other components such as sulfite (chemical oxygen demand -- COD) plus any remaining surface-moisture constituents.

- Cold winters may severely restrict dragline operations during cold-weather months and jeopardize continuity of FGD operation. In addition, freeze-thaw action is of concern so far as maintaining the integrity of dikes.

LIQUID EFFLUENT

In view of the low surface-moisture content of coarse-grained solid waste from FO-FGD systems, provisions for management of saline FGD liquor blowdown are essential for pollution control. Thus, the FO-FGD system designer would be expected to seek all practical design means to manage waste effluent such that acceptably low scrubber-liquor salinity is maintained (4).

Significance of Discharge

Although FO-FGD blowdown may, by volume, be only a small proportion of the total power-plant wastewater, it contains dissolved-salts concentrations approaching those in sea water and thus may exceed the assimilative capacity of available receiving streams. Such outfall, untreated, can also contain objectionable concentrations of trace metals such as arsenic, mercury, cadmium and lead that may originate from the fired coal (5).

It should be noted that liquid purge outfall such as required from FO-FGD is not always necessary in a natural oxidation mode because these latter systems yield a solid waste (primarily calcium sulfite) that is fine-grained, thixotropic and highly water retentive (containing approximately 50% moisture). The encapsulation of a substantial amount of FGD liquor in the discarded waste mass may provide the liquid purge necessary to maintain the scrubber dissolved salts at an acceptable concentration. However, natural oxidation sludges have many disadvantages as compared to those produced in a forced oxidation mode. These disadvantages include both physical properties –– the completed disposal site cannot be reclaimed for uses requiring the soil to support structural loads –– and chemical properties. The principal contaminants in the sulfite sludge are dissolved salts (principally sulfates), chemical oxygen demand due to sulfites, and soluble heavy-metal trace elements originating in the coal. Thus leaching and runoff from sites used for disposal of such solids have also been of significant environmental concern. An important remedy is sludge fixation (stabilization), which consists of chemical processing of scrubber sludge to facilitate handling, transportation, placement, and consolidation at the ultimate disposal site. One of the methods used is pug-mill blending and aging of mixtures of FGD filter cake, dry fly ash and other dry additives, followed by landfill compaction. An alternative procedure involves the addition of silicate-containing materials to thickened FGD slurry before discharge to a pond site. Extensive testing indicates that these methods of chemical treatment significantly improve the load-bearing characteristics of FGD sludge, reduce the solubility of the major chemical species by a factor of 2 to 4 and reduce sludge permeability by an order of magnitude or more (6). Further, as outlined previously, the encapsulation of a substantial amount of FGD liquor in the consolidated waste mass from natural oxidation FGD provides the option for all-in-one, combined disposal of all solid and liquid waste without discharge of a liquid outfall.

Alternatives for Treatment/Ultimate-Disposal

The liquid purge may be treated to precipitate trace-metal constituents and reduce their concentration to an acceptable level so that the effluent stream may be discharged to a suitable receiving stream or water body. Such treatment followed by outfall discharge is the economically preferred method of FGD liquor disposal when it can be permitted.

Candidate wastewater treatment methods to reduce trace metal concentrations include:

- Chemical precipitation
- Ion exchange (exchange of complex anionic species)
- Sulfide precipitation
- Chelation/absorption.

As indicated above, because of the large dissolved solids concentration of purge-liquor, available surface water receiving streams at inland sites may be unable to tolerate the environmental impact of saline outfall discharge of this type. Blowdown liquor after treatment can, in this case, be concentrated utilizing commercially available equipment to crystallize dissolved salts to a comparatively small mass of salt cake that can be discarded at a secure landfill disposal site or used as a rock salt byproduct.

UTILIZATION OF FGD GYPSUM

INTRODUCTION

Annual consumption of crude gypsum in the United States is estimated to be approximately 20 million tons, with 30 - 40 percent being imported, mainly from Canada and Mexico. The average value of crude gypsum was estimated to be $9.00 per ton (F.O.B. mine) in 1982 (7).

The sale and utilization of gypsum from FGD would help to offset a small portion of the operating cost of the FGD system and would minimize the environmental impact of and use of land for ultimate disposal of FGD solid wastes. There are several factors to be considered in determining the feasibility of FGD gypsum sale and utilization. These include technical and economic aspects as well as incentives, problems and barriers which are specific to the circumstances in the United States.

TECHNICAL FEASIBILITY

The gypsum formed by FO-FGD is unlike natural gypsum, the major differences being: bulk handling properties (moisture content, grain size), quantity of impurities, and nature of impurities. Natural gypsum is mined as a coarse, low-moisture rock at approximately 70 - 90 percent purity, with limestone and/or anhydrite (non-hydrated calcium sulfate) comprising the majority of impurities. FGD gypsum is produced as a higher moisture filter cake of approximately 90-95 percent purity, primarily coarse-grained crystals along with ash, lime or limestone components and dissolved salts comprising the majority of impurities. The technology to make and use board-grade and cement-grade FGD gypsum exists and is currently in use in Japan and Europe. Depending on end use (cement, prefabricated products or agriculture), the FGD gypsum may require processing (beneficiation) before it is suitable for use. For example, gypsum is used in cement manufacture both as a grinding aid and as a source of calcium sulfate to retard setting. To be acceptable for this use, FGD gypsum cake would need only minimal washing (to ensure chloride content of less than 1 percent), but requires drying/agglomeration to achieve suitable grinding properties. To be used in gypsum board manufacture, FGD gypsum cake requires washing, (the amount depending on chloride content, unwashed, and board manufacturers' specification) and may require drying/agglomeration. Additionally, suitable FGD process design and control is vital to ensure an adequate degree of oxidation of sulfite.

ECONOMIC FEASIBILITY

There will be costs incurred by the utility in converting the system design and ultimate operation from disposal to utilization of FGD gypsum. The net cost to the utility will be the beneficiation costs less the hypothetical disposal cost if the gypsum was not to be used. However, the utility should be able to sell the gypsum, which may either offset the production cost or result in a net credit to the utility. As described earlier, natural gypsum has an average value of $9.00 per ton (F.O.B mine). Thus the apparent value of gypsum to the user is typically $9.00 per ton plus transportation costs. Transportation costs for gypsum imported from Nova Scotia and delivered to the southeastern U.S.A. were estimated to be $15.00 per ton in 1982 (8). Thus the value of gypsum to a user can range from $9.00 to $20 - 30 per ton at site of use, the F.O.B. value in Nova Scotia being as high as $15.00 per ton. However, in the United States, a depletion allowance of 14 percent of the price of gypsum is granted to mining companies as a tax benefit to compensate for the depletion of mine reserves (9). This depletion allowance is applied to the price of gypsum F.O.B. point of sale, not minesite, and therefore includes transportation costs (plus other costs for handling and preparation). Thus, the tax credit to the gypsum company effectively reduces the cost of mined gypsum. Although it is beyond the scope of this study to quantify the benefits of the depletion allowance, it is suggested that the actual cost of gypsum to a gypsum mining company/board-manufacturer is substantially less than this apparent value of $9.00 to $20 - 30 per ton at site of use.

The costs to the utility of producing usable gypsum are estimated to range from $5 to $18 per ton for board-grade gypsum and from $8 to $11 per ton for cement-grade gypsum (8). The wide range of costs is due to the diversity of possible methods of processing the gypsum cake (10). This is in turn due to the wide variance in board manufacturers' specifications (see Table 2) and the various options available for dewatering (rotary drum filter, horizontal belt filter, etc.), agglomeration (pelletization, extrusion, briquetting) and drying (fuel-fired vs regenerative). However, the cost of producing usable gypsum may be partially or completely offset by hypothetical disposal costs. The comprehensive analysis of the costs associated with landfilling a filtered gypsum waste is a complex procedure. A preliminary assessment of these costs yields disposal costs ranging from $10.00 to $25.00 per ton (11). Costs associated with such disposal include land costs, site preparation costs, dewatering costs, transport to disposal site, wastewater treatment and reclamation. The wide range of costs is due to variance in land availability and suitability, and uncertainty in wastewater treatment costs. The estimate of wastewater treatment cost is uncertain as it depends on the quantification of the flow volume of liquid purge required to limit the dissolved solids in the scrubber slurry to an acceptable level. This limit and volume depend on coal chloride content, washed-cake quality criteria, materials of construction, etc.

Table 2. WALLBOARD

MANUFACTURERS' SPECIFICATIONS

FOR FGD GYPSUM FOR USE IN BOARD

	National Gypsum	Georgia Pacific	U.S. Gypsum
Gypsum Content, %, Min.	94	90	95
Calcium Sulfite Hemihydrate, %, Max.	0.5	N.S.	2
Na, ppm, Max.	500	200	75
Cl, ppm, Max.	800	200	120
Mg, ppm, Max.	500	N.S.	50
Free Water, %, Max.	1	12	10
pH	6-8	3-9	6.5-8
Particle size, Min.	xy= 2000 sq. micron	N.S.	20 (mean) micron

Note: N.S. = Not specified

In summary, the cost of FGD gypsum beneficiation is offset by the disposal costs which would have been incurred in a non-utilization landfill disposal option. If lower-cost wet stacking is the nominal disposal option, gypsum beneficiation may be seen to cost more than disposal. However, because the gypsum can be sold for use in gypsum board or cement, a net credit may be realized if arrangements are made for a suitable user. If this is so, the economics of production and sale of by-product gypsum can be expected to be favorable.

PROBLEMS AND BARRIERS

Many influences within existing gypsum markets affect the potential for recovery of gypsum from by-product sources. In each case, many factors (location, capacity, quality, etc.) will significantly affect the possibility of such by-product gypsum utilization. Assessment of these factors is complex since there has been no extensive by-product gypsum utilization in North America and because of the vertical integration of major firms in the gypsum industry (i.e. gypsum mines, transportation systems and board manufacturing plants are generally all owned and operated by gypsum companies).

Principal obstacles to gypsum use by board and cement manufacturers are as follows:

- Natural gypsum is a plentiful, inexpensive, raw material. (Ease of transportation of by-product to the potential user(s) as contrasted with the conventional source is a crucial factor in its economic use. As with natural gypsum, the transportation cost will generally be the most significant part of the delivered cost.)
- Gypsum users are reluctant to convert to by-product gypsum unless both quality and quantity of the gypsum are assured. Given the fact that electric utilities exist to produce power and are not accustomed to serving markets outside the energy sector, warranting of by-product quality and quantity may be a significant problem.
- When ample land areas for disposal are available close to the power plant, the impetus for producing salable gypsum may not be present.
- Historically, the gypsum and utilities industries have each been very conservative and insular. The production and use of by-product gypsum by these industries will require policy changes that they may find difficult to implement.

ACTIVE BY-PRODUCT FGD PROJECTS IN NORTH AMERICA

The four plants (Table 1) that have committed to date to the production of salable gypsum vary so far as geographic location, system size and type of utility company, but all are similar so far as:

- The basis for use of forced oxidation including the impetus for production of usable gypsum
- System design and operation
- The prospect of gaining use of the by-product.

KEY MOTIVATING FACTORS

Aside from major process benefits of forced-oxidation mode operation, including improved control of internal scaling and increased reagent utilization, by-product gypsum applications to date have incorporated forced-oxidation with production of a salable product because of substantial constraints preventing use of an extensive area of nearby land for throwaway type disposal. Also, apart from potential sale/use of the gypsum, this manner of applying forced-oxidation ensures that the SO_2-catch is rendered in a cake form that may readily be handled and temporarily stockpiled, and thereafter ultimately transported any necessary distance to an off-the-site point or points of delivery. Under such circumstances, and because of the comparative ease and limited cost to adjust the design and quality of the solid waste to yield a usable product, each of the projects has with adequate incentive (but without agreements or contracts for product sales), committed to a design compatible with by-product utilization. The resulting gypsum yield is a benign and highly acceptable solid waste except for its significant solubility in water, and can possible be sold to existing users or suppliers of natural gypsum located reasonably close to the power plant.

DESIGN AND OPERATION

In the FGD systems sold for usable gypsum production, the bulk of SO_2 removal is achieved in counterflow absorbers that discharge slurry effluent to gas quenchers (precoolers) that yield an exit scrubbing slurry having a suitably low percent calcium sulfite in the solid phase. No isolation of the gypsum product from hydrogen chloride gas or other non-SO_2 components of the raw flue gas removed in wet collectors is achieved in the gas scrubbing section. After dewatering of this final slurry, a wet cake product containing as little as 8 percent surface moisture is discharged from rotary vacuum filters equipped with cake washing accessories. The concentration of objectionable water-soluble constituents in the gypsum cake product is limited both by limiting of concentrations in the scrubber liquor through discharge of liquid purge from the slurry dewatering system to an acceptable receiver, and by the washing of the cake after it forms on the filter drum. The concentration of objectionable solid-phase constituents other than calcium sulfite is limited by use of a high-efficiency electrostatic precipitator upstream of FGD, and of limestone reagent of adequate quality/purity.

POTENTIAL FOR USE OF GYPSUM PRODUCED

These FGD systems are capable of continuous production of gypsum that meets specifications for intended use, wallboard in the case of three plants in Table 1 and other less critical uses in the case of the relatively small output of Muscatine Power and Water. Market demand for natural gypsum in the regions in which the plants are located is sufficient to assimilate the FGD by-product yields, but each of the FGD systems has been purchased without benefit of a gypsum sales agreement with a potential user(s), and no such contracts have been announced to date.

Advantages of By-Product Substitution

The principal advantages to potential buyers of use of these FGD gypsum outputs are:

- The material may be expected to be supplied at a uniformly acceptable level of quality.
- The quantity of supply is substantial in relation to the contemplated end use(s).
- Particularly in Florida where natural gypsum is presently shipped long-distance from Nova Scotia to supply wallboard manufacturers, or principally from Spain in the case of gypsum rock for cement set retarder, substantial savings in transportation cost may be realized.

Means for Meeting User Requirements

Provisions for drying and/or agglomerating of the FGD gypsum may in some instances be necessary since prospective users may incur excessive costs or inconvenience using raw washed FGD cake instead of gypsum rock. Presence of surface moisture may also make it difficult to handle and store the material in existing facilities designed for natural gypsum rock. Moreover, water adds to shipping weight and unit transportation cost and requires added energy use by the buyer since drying will generally be needed for purposes of the end use. In specific instances, for example supply of gypsum for use as cement set retarder if relevant, agglomeration of the product by briquetting, extruding or other suitable means may be essential for successful marketing of plant output.

Principal Obstacles to Use of By-Product

Lack of success in sales contract negotiations to date points to the need for an equitable, generally acceptable, and broadly applicable basis for pricing of the product. This may be essential in establishing long-term by-product gypsum sales contracts between power plants and the major gypsum users best able to benefit from use of the new source of supply. The depletion allowance on natural gypsum authorized by the U.S. Internal Revenue Service Code-613B significantly reduces the financial benefit to the vertically integrated, North American gypsum industry of substitution of by-product. If the allowance is to continue to apply in an era of newly and plentifully available by-product gypsum suitable for principal market uses, the by-product producer must realistically evaluate the supply economics and competitive pricing of the product taking into account the purchasing disincentive created by the IRS tax allowance.

CONCLUSIONS

1. Forced oxidation type flue gas desulfurization (FO-FGD) technology offers a design means for forming a usable gypsum by-product instead of a throwaway solid waste.

2. Widespread application of FO-FGD in U.S.A. since 1978 has been almost entirely in a throwaway waste mode with little apparent incentive for usable gypsum production. Only in limited instances in which the throwaway-disposal option is very constrained, i.e., 13 percent of all U.S. FO-FGD capacity purchased to date, is the FGD design predicated on manufacture of usable gypsum, and in none of these cases, four in number, has a formal commitment to purchase or use the by-product been obtained to date.

3. Cost advantages point to possible future use of "wet stacking" type disposal in FO-FGD design and to significant environmental impacts that are largely circumvented in earlier natural oxidation type FGD designs that employ sludge fixation by use of fly ash and other additives.

4. Available technology and the successful widespread utilization of FGD gypsum in Japan and West Germany indicate that its use is technically feasible.

5. It is likely that, with or without beneficiation, the sale of FGD gypsum to the board and cement manufacturers would be economically preferable for the utility company in lieu of the throwaway-disposal alternative.

6. Institutional barriers to the use of FGD gypsum throughout North America include the high degree of vertical integration of the gypsum wallboard manufacturing industry, the disincentive for by-product gypsum utilization created by the gypsum depletion allowance under U.S. Internal Revenue Service income tax codes, and the abundance at diverse locations of natural gypsum of adequate quality.

7. Additional major factors influencing the profitability of production and sale of usable FGD gypsum include the magnitude of alternative hypothetical throwaway-disposal costs, transportation costs to user(s), and costs of transportation of user's nominal source(s) of natural gypsum supply.

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OPERATING EXPERIENCE WITH THE CHIYODA THOROUGHBRED 121 FLUE GAS DESULFURIZATION SYSTEM

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ABSTRACT

This paper reviews the design and operating experience of the Chiyoda THOROUGHBRED 121 Flue Gas Desulfurization System installed and operated by Mitsubishi Petrochemical Company at their Yokkaichi, Japan, complex.

The plant consists of a single scrubber and treats 260,000 Nm^3/h (162,000 scfm) of flue gas from a 280 T/h (87 MW equivalent) boiler burning high sulfur oil. Plant operation, since startup in May 1982, has been smooth and trouble free over a wide range of operating conditions. Plant reliability has been 100%. The plant is operated at a SO_2 removal efficiency of 97 to 99 percent for inlet SO_2 concentrations ranging between 1000 and 2000 ppm. Limestone utilization is greater than 99 percent and the dry, gypsum by-product is sold to a wallboard manufacturer. Operating and maintenance functions for the system are minimal and completely absorbed by normal boiler plant operations.

INTRODUCTION

In March of 1981, the Mitsubishi Petrochemical Company awarded Chiyoda Chemical Engineering and Construction Company, Ltd. a contract to install its patented, second generation Chiyoda THOROUGHBRED 121 (CT-121) Limestone/Gypsum System. The System was selected due to its low space requirement, simple operation, and low plant cost. In addition, Chiyoda's engineering capability and expertise were highly regarded based on Mitsubishi's experience with two first generation CT-101 Systems.

The CT-121 System was to be designed to remove 97% of the SO_2 contained in the flue gas from a new boiler being constructed by Mitsubishi Petrochemical Company at its Yokkaichi complex. The boiler provides steam for generating 40 MW of electricity and 47 MW equivalent of process steam. Fuel oil containing 3 to 4% sulfur is burned, producing a flue gas SO_2 concentration of 1000 to 2000 ppm.

In addition to removal of SO_2, the System was also designed to remove particulate matter (mostly carbon dust) from the boiler, thereby taking advantage of the high particulate removal capability of the CT-121 Process and eliminating the need for installing an ESP. The CT-121 System described herein, therefore, incorporates equipment such as a fly ash thickener and filter press that ordinarily would not be required.

The CT-121 System, including particulate removal equipment, was routinely started up on May 11, 1982. The System has met or exceeded all performance guarantees and operation has been smooth and trouble free over a wide range of boiler loads and inlet SO_2 concentrations.

GENERAL SYSTEM DESCRIPTION

Table 1 is a summary of the major design conditions for the CT-121 FGD and particulate removal system and Table 2 is an equipment list including size and materials of construction.

TABLE 1. SUMMARY OF DESIGN CONDITIONS OF THE MITSUBISHI PETROCHEMICAL CT-121 PLANT

Flue Gas Source	280,000 kg/h (617,000 lb/h) oil-fired boiler (87 MW equivalent)
Flue Gas Flow Rate	260,000 Nm^3/h (162,000 scfm)
Flue Gas Temperature	140-160°C (284-320°F)
SO_2 Inlet Concentration	1500-2000 ppm
Particulate Loading	200 mg/dNm^3 (0.09 gr/dscf)
SO_2 Removal Efficiency	97%
Particulate Emission	< 50 mg/dNm^3 (0.02 gr/dscf)
Reagent	Powdered limestone
By-Product	Saleable gypsum
Total Plot Area*	1540 m^2 (0.4 acres)
Schedule	
Start Construction	June 1981
Start Operation	May 1982

*Includes gypsum by-product storage, particulate removal equipment and 120 meter solid FRP wet stack.

TABLE 2. EQUIPMENT LIST

Equipment	Type	Size	Material	Remarks
		CT-121 FGD EQUIPMENT		
Flue Gas Fan	Turbo	260,000 Nm^3/h (162,000 scfm)	c.s.	
Precooler	Cylindrical		c.s. with FlRP lining	
Jet Bubbling Reactor	Cylindrical	9.8 mØx10 mH (32'Øx33'H)	316L s.s. FRP	with Agitator
Air Blower	Turbo	5,000 Nm^3/h (3100 scfm)	cast aluminum	
Overflow Tank	Cylindrical		c.s. with FlRP lining	with Agitator
Limestone Silo	Cylindrical	6.8 mØx19.6 mH (23'Øx65'H)	c.s.	
Limestone Slurry Tank	Cylindrical		c.s. with FlRP lining	with Agitator
Limestone Feeder	Screw Feeder	2,300 kg/h (5100 lb/hr)	c.s.	
Mother Liquor Tank	Cylindrical		c.s. with FlRP lining	
Centrifuge	Screw Decanter	3 x 1,330 kg/h (3000 lb/hr)	316L s.s.	
Mist Eliminator	Chevron		c.s. with FlRP lining, P.P.	
Gas Cooling Pump	Centrifugal		c.s. with Rubber lining	
Gypsum Slurry Pump	Centrifugal		c.s. with Rubber lining	
Overflow Pump	Centrifugal		c.s. with Rubber lining	
Limestone Slurry Pump	Centrifugal		c.s. with Rubber lining	
Mother Liquor Pump	Centrifugal		c.s. with Rubber lining	
		FLY ASH REMOVAL EQUIPMENT		
Fly Ash Thickener	Cylindrical		c.s. with FlRP lining	with Rake
Neutralization Tank	Cylindrical		c.s. with FlRP lining	with Agitator
Filter Press	Filter Press		c.s. with Rubber lining, 316L s.s.	
Filtrate Tank	Cylindrical		c.s. with FlRP lining	
Thickener Overflow Pump	Centrifugal		c.s. with Rubber lining	
Thickener Underflow Pump	Centrifugal		c.s. with Rubber lining	
Filter Feed Pump	Centrifugal		c.s. with Rubber lining	
Filtrate Pump	Centrifugal		c.s. with Rubber lining	
Blow Down Tank	Cylindrical		c.s. with FlRP lining	
Blow Down Pump	Centrifugal		c.s. with Rubber lining	

Figure 1 is a plot plan of the entire system showing the arrangement of major equipment including the stack. Figure 2 is an overview of the FGD portion of the System.

Flue gas to the System fluctuates daily between 50-100% of design capacity due to changes in boiler load. The SO_2 concentration also fluctuates widely due to the variety of fuel oils burned.

The control of the power plant, particulate and SO_2 removal equipment is centralized and computer controlled. Both the boiler and CT-121 System can be started up, operated, and shutdown from the central control room.

The heart of the CT-121 Process is the patented absorber which is called a Jet Bubbling Reactor (JBR). This device, illustrated in Figure 3, allows SO_2 to be absorbed, neutralized, oxidized, and crystallized in a single vessel. The overall reaction equation for the process is:

$$SO_2 + CaCO_3 + \frac{1}{2}O_2 + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O + CO_2$$

PROCESS DESCRIPTION

Figure 4 shows the process flow for the CT-121 FGD System including particulate removal equipment. Flue gas from the air preheater is pressurized by the Flue Gas Fan and introduced into the Precooler. The gas is contacted with a fine spray of recirculated water to humidify and cool the gas to its adiabatic saturation temperature. The Precooler is also designed to remove particulates and other impurities contained in the flue gas. The humidified gas flows to the JBR where it is bubbled into a shallow layer of absorbent. SO_2 is absorbed and precipitated as calcium sulfate (gypsum) by the addition of limestone slurry and oxidizing air.

The desulfurized gas leaving the JBR flows through a two stage mist eliminator to remove carryover mist. The mist eliminator is washed once per day with mother liquor to remove accumulated fine gypsum crystals.

The level in the JBR is held constant by maintaining a liquid flow over a weir to the overflow tank. Overflow liquid is recycled to the JBR.

Gypsum slurry is withdrawn from the JBR and is mechanically dewatered using a solid bowl decanter centrifuge. The by-product gypsum is essentially a dry material and is collected in a temporary storage area directly below the centrifuge. Mother liquor from the centrifuge flows to the mother liquor tank and recycled to the System.

Powdered limestone from the silo is fed to the limestone slurry tank via a screw feeder and mixed with mother liquor to obtain a 20 wt.% slurry. The slurry is pumped to the JBR on flow control based on a feedforward boiler load signal and a feedback JBR pH signal.

A portion of the Precooler cooling water flows to a fly ash thickener to concentrate captured particulates. The overflow liquid is returned to the Precooler and the underflow is pumped to a neutralization tank. A small amount

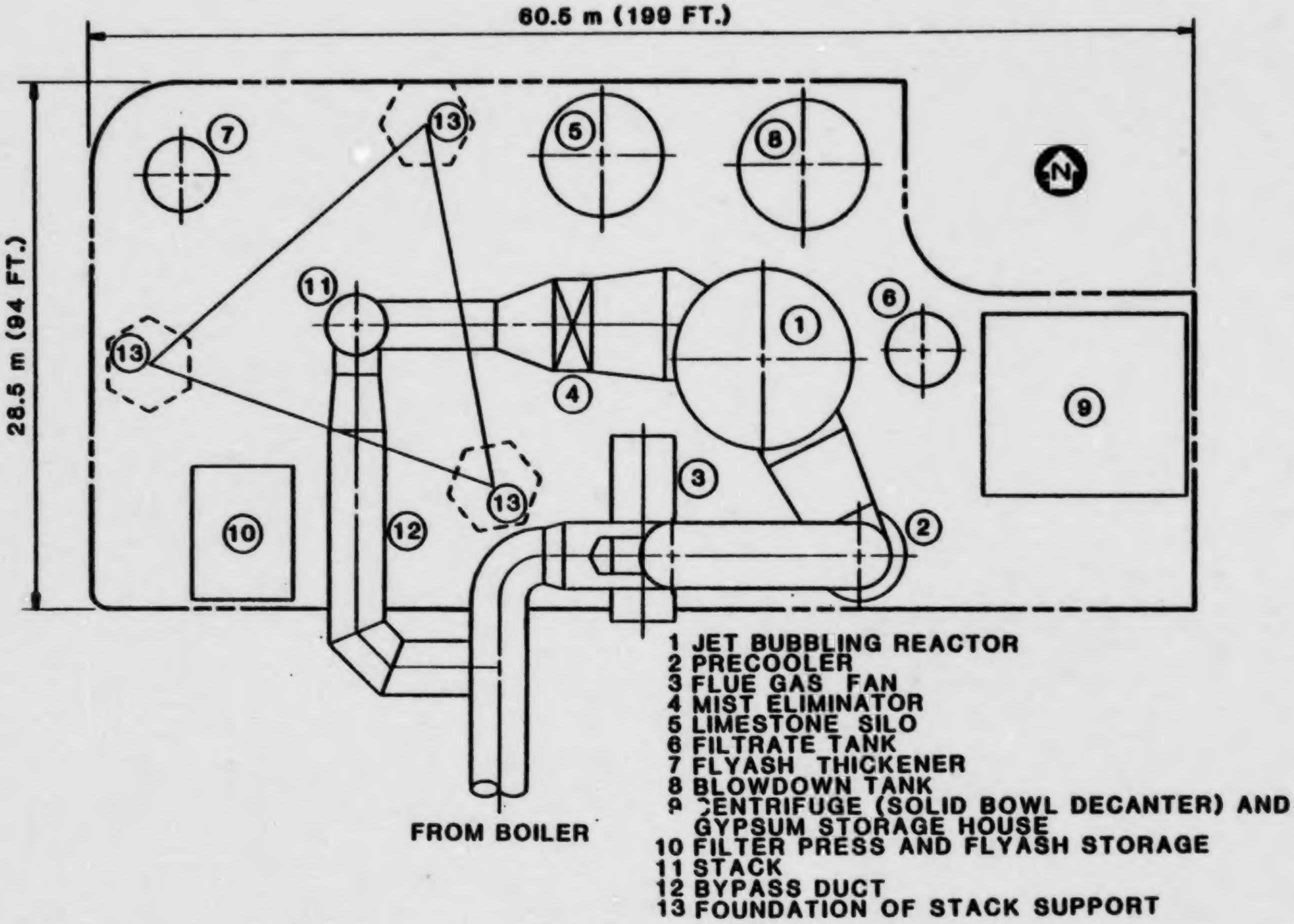
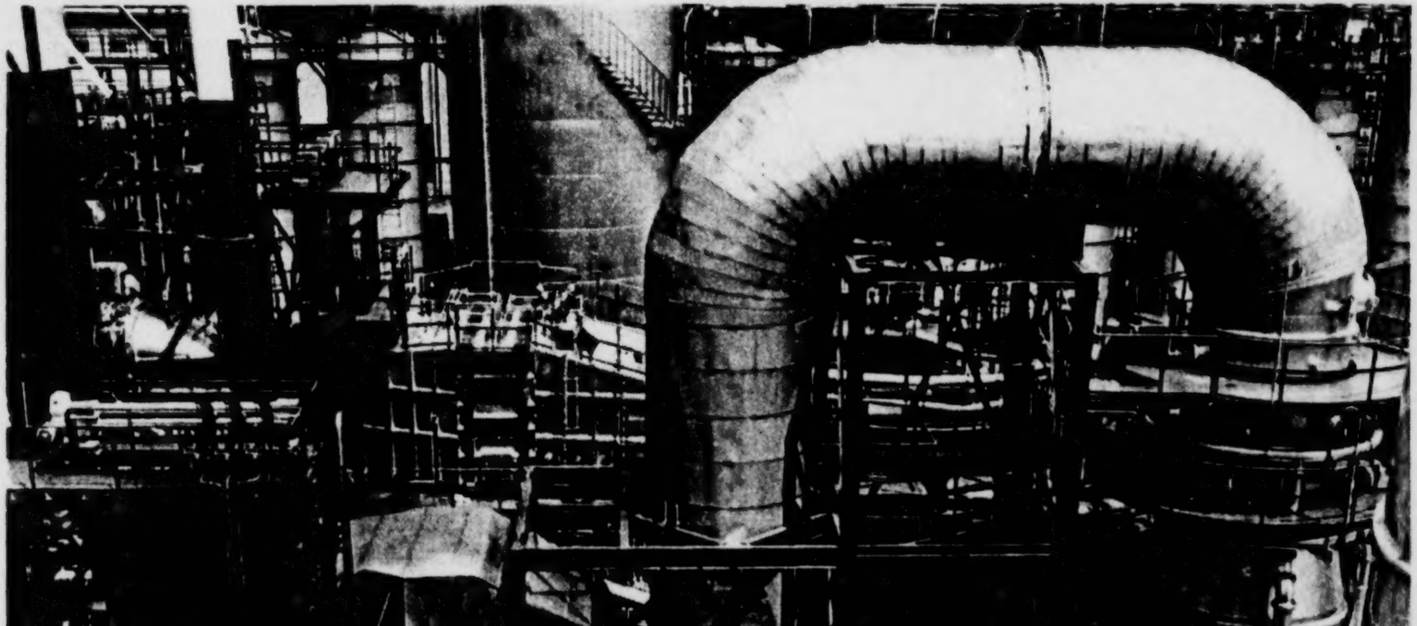


FIGURE 1. ARRANGEMENT OF THE CT-121 FGD PLANT FOR MITSUBISHI PETROCHEMICAL CO.



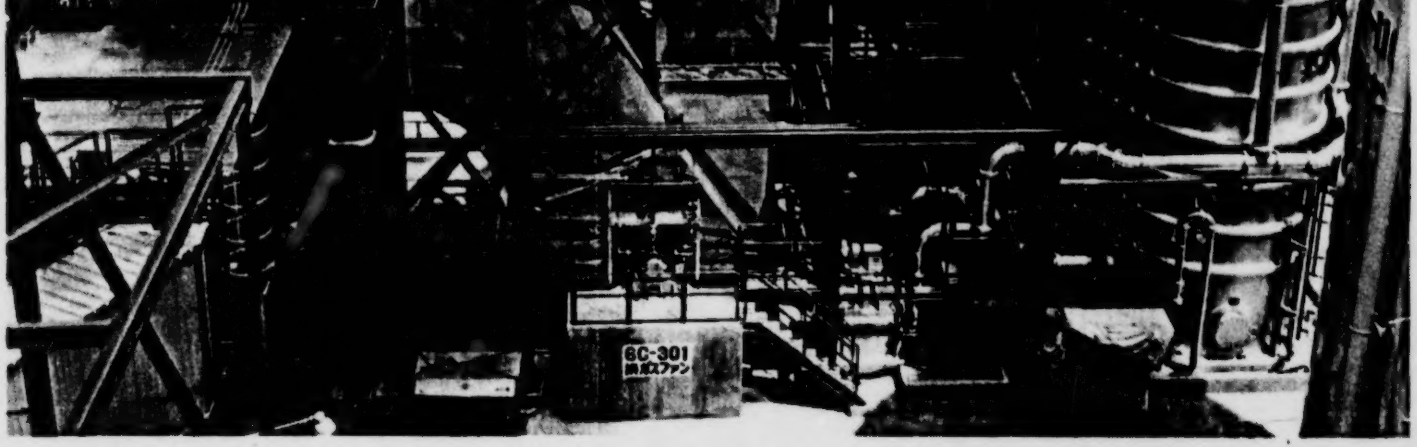


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Figure 2. OVERVIEW OF THE MITSUBISHI PETROCHEMICAL CT-121 PLANT

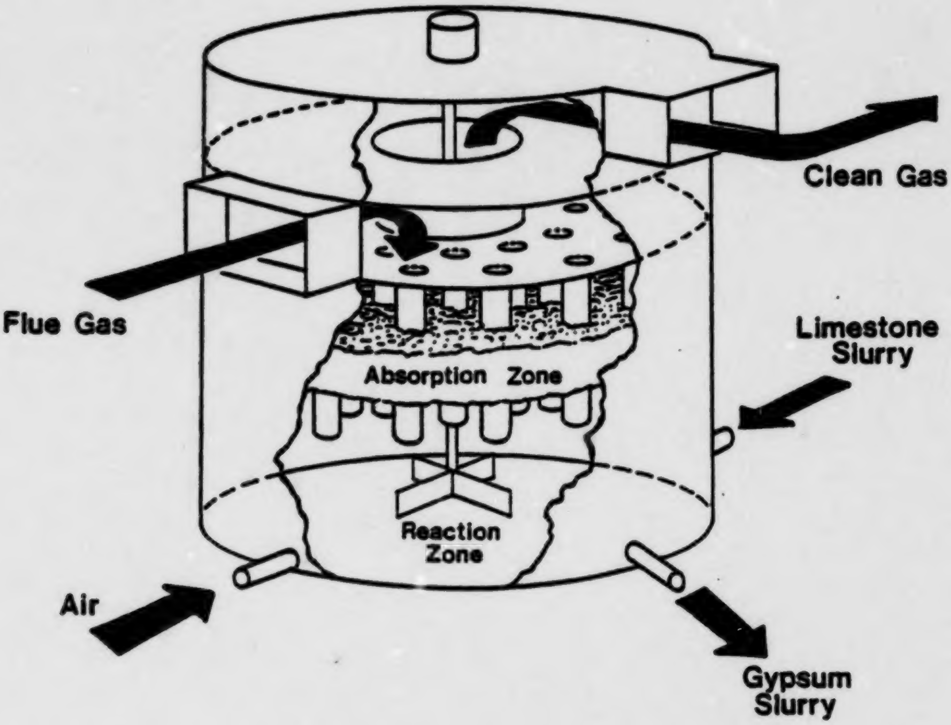


Figure 3. Jet Bubbling Reactor

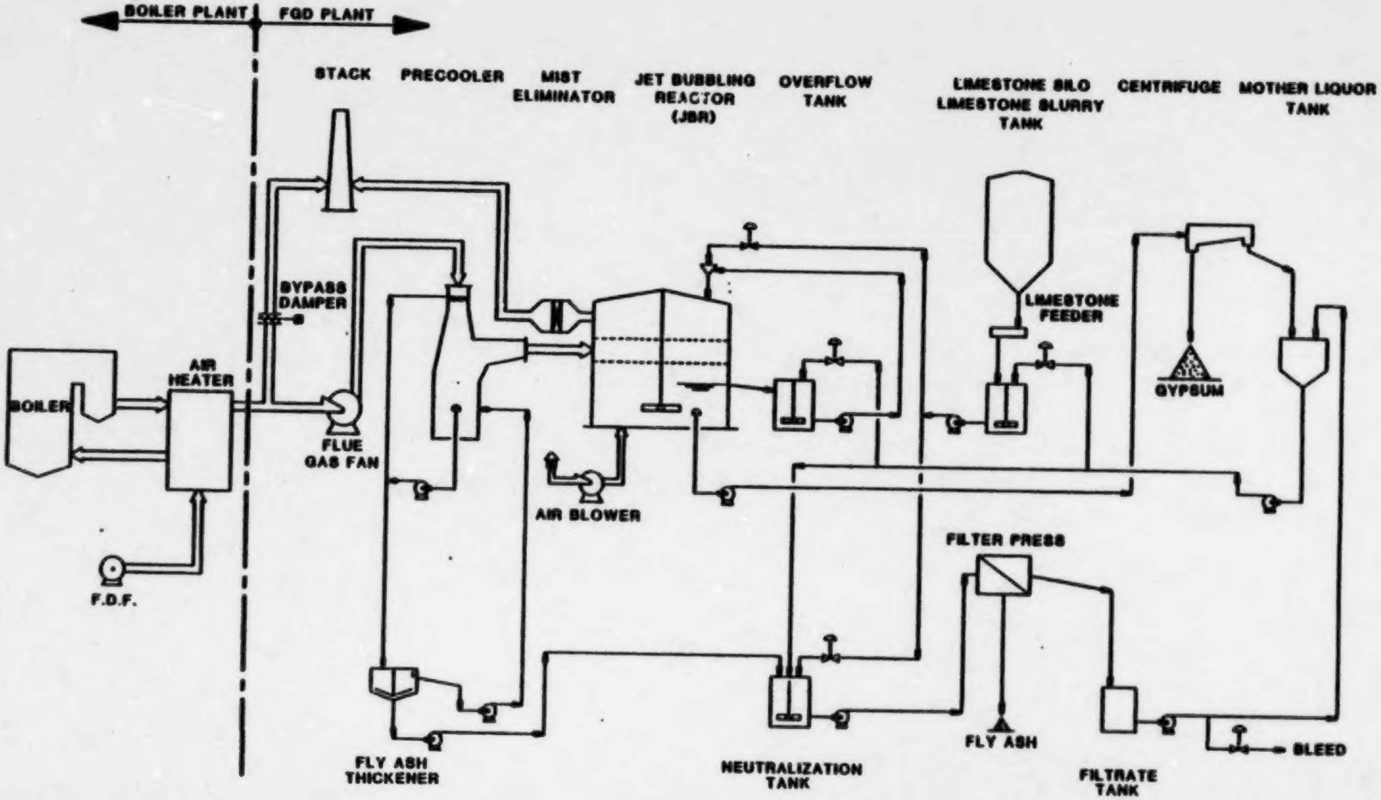


FIGURE 4. PROCESS FLOW SHEET OF THE CT-121 FGD PLANT FOR MITSUBISHI PETROCHEMICAL CO.

of mother liquor is also added to this tank. The combined stream is neutralized with limestone slurry. The neutralized liquor, containing approximately 2 wt.% solids, is pumped to a filter press where the solids are removed. The filtrate is collected in a filtrate tank and recycled to the system via the mother liquor tank. A small amount of the filtrate (1 to 3 m^3/hr (4 to 13 gpm)) is bled from the filtrate tank to maintain system water balance.

OPERATING RESULTS

PLANT OPERATION

The CT-121 System was smoothly and routinely started up on May 11, 1982. Tables 3a and 3b are summaries of the operating history of both the boiler and the CT-121 System. As shown in Table 3b, System reliability has been 100%. This is particularly impressive considering the System consists of only a single scrubbing module and is operated under severe conditions.

DESULFURIZATION EFFICIENCY

The SO_2 removal efficiency, as shown in Table 3a, has averaged 97 - 99% for inlet SO_2 loadings ranging between 1000 - 2000 ppm.

Figures 5 and 6 show the effect of boiler load and SO_2 inlet concentration changes on desulfurization efficiency. These two Figures illustrate the CT-121 Process' capability to maintain stable desulfurization under widely fluctuating boiler loads and inlet SO_2 concentrations.

OPERABILITY AND CONTROLLABILITY

The boiler load changes daily from 50 - 100% of design capacity. The System responds well and automatically to such load fluctuations. Figure 7 illustrates the CT-121 System response to a typical change in load.

The CT-121 System is extremely simple to operate and requires little operator attention. System operation is basically automatic, being controlled by a digital computer control system.

The flow of flue gas to the System is set by boiler load. Flow is automatically adjusted by a pressure controller installed on the suction side of the Flue Gas Fan which controls the position of the fan inlet vanes. A bypass damper is installed to protect the boiler against sudden changes in draft.

The limestone slurry concentration is maintained constant by adding limestone in proportion to mother liquor feed by using a proportional control loop. Limestone slurry is monitored using an in-line density meter.

The gypsum concentration in the JBR is maintained at about 15 wt.% by pumping a small slip stream of slurry to the centrifuge. The flow rate is controlled in proportion to the SO_2 load signal which is the product of the boiler load signal and the SO_2 inlet concentration.

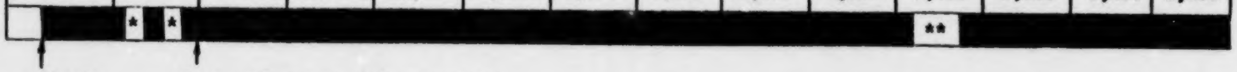
TABLE 3a. OPERATING HISTORY OF THE MITSUBISHI PETROCHEMICAL, CT-121 FGD PLANT

Operating Conditions

Flue gas source	oil-fired boiler
Boiler capacity	280 T/h steam (87 MW equivalent)
Startup date	May 11, 1982
Gas flow rate	100,000-246,000 Nm^3/h (62,000-153,000 scfm)
Gas temperature	140-160°C (284-320°F)
SO_2 concentration	700-2,300 ppm
Desulfurization efficiency	96-99%

Operating Summary

1982								1983					
MAY	JUN	JUL	AUG	SEP	OCT	NOV	DEC	JAN	FEB	MAR	APR	MAY	JUN
96[a] 1,500[b]	96 1,500	98 1,000	99 1,000	98.5 1,000	98.5 1,500	97.5 2,000	97.5 2,100	98 2,000	98 1,800	98 1,800	97.5 1,600	98 2,000	98 2,000



Startup
5/11/82

Start of commercial operation

Cause of Shutdowns

■	Operating	*Scheduled shutdown for boiler/turbine inspection
□	Shutdown	**Scheduled shutdown for first annual inspection of steam generating plant.

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[a] Average monthly SO_2 removal, %

[b] Average monthly inlet SO_2 conc., ppm

TABLE 3b. CT-121 RELIABILITY PARAMETERS (MAY 11, 1982 THROUGH JUNE 30, 1983) FOR THE MITSUBISHI PETROCHEMICAL FGD PLANT

Parameter	Value (%)	Total Time (hours/hours)
Reliability [a]	100.0	9,249/9,249
Availability [b]	98.1	9,816/10,008
Operability [c]	99.9	9,249/9,255
Utilization Factor [d]	92.4	9,249/10,008

[a] Reliability - hours the FGD System was operated divided by the hours the FGD system was called upon to operate.

[b] Availability - hours the FGD System was available for operation (whether operated or not), divided by the hours in the period.

[c] Operability - hours the FGD System was operated divided by the boiler operating hours in the period.

[d] Utilization Factor - hours that the FGD System was operated divided by the hours in the period.

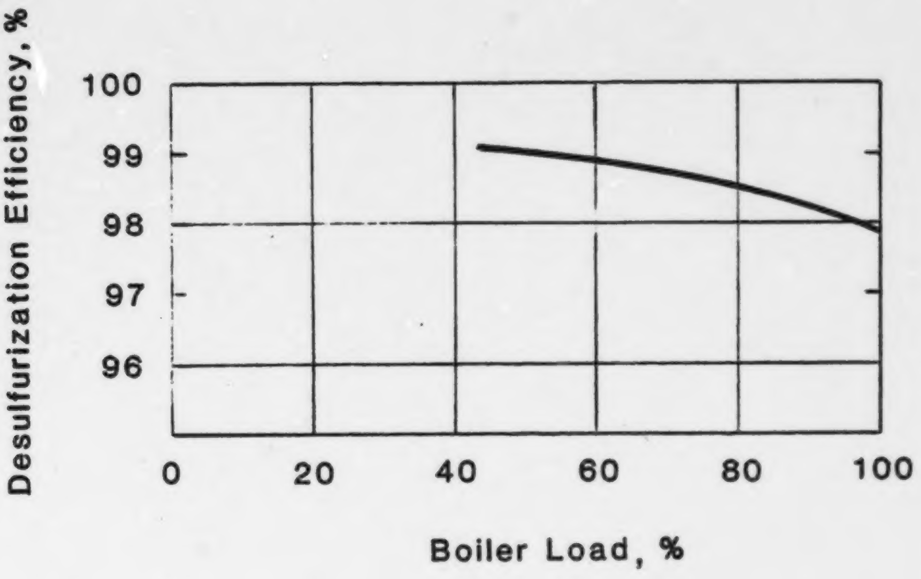

Figure 5. EFFECT OF BOILER LOAD ON DESULFURIZATION EFFICIENCY

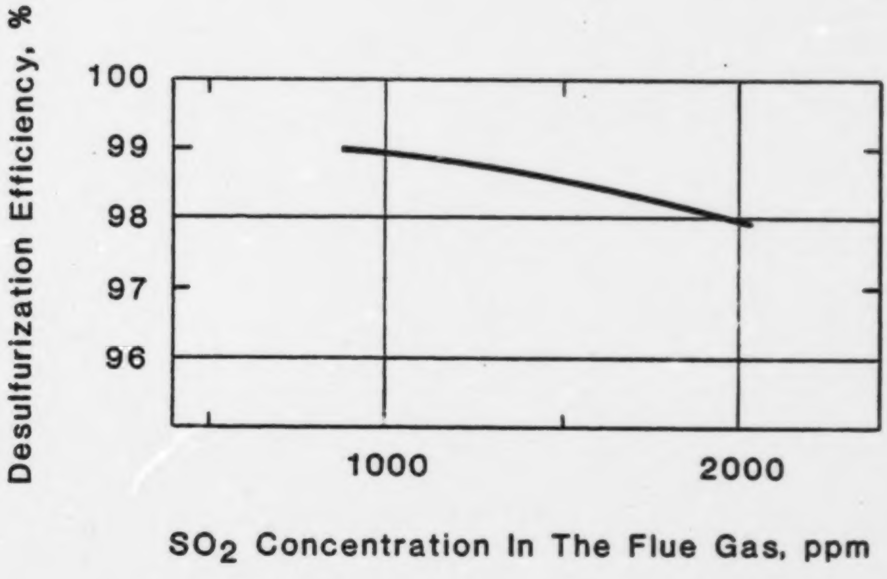

Figure 6. EFFECT OF SO_2 CONCENTRATION ON DESULFURIZATION EFFICIENCY

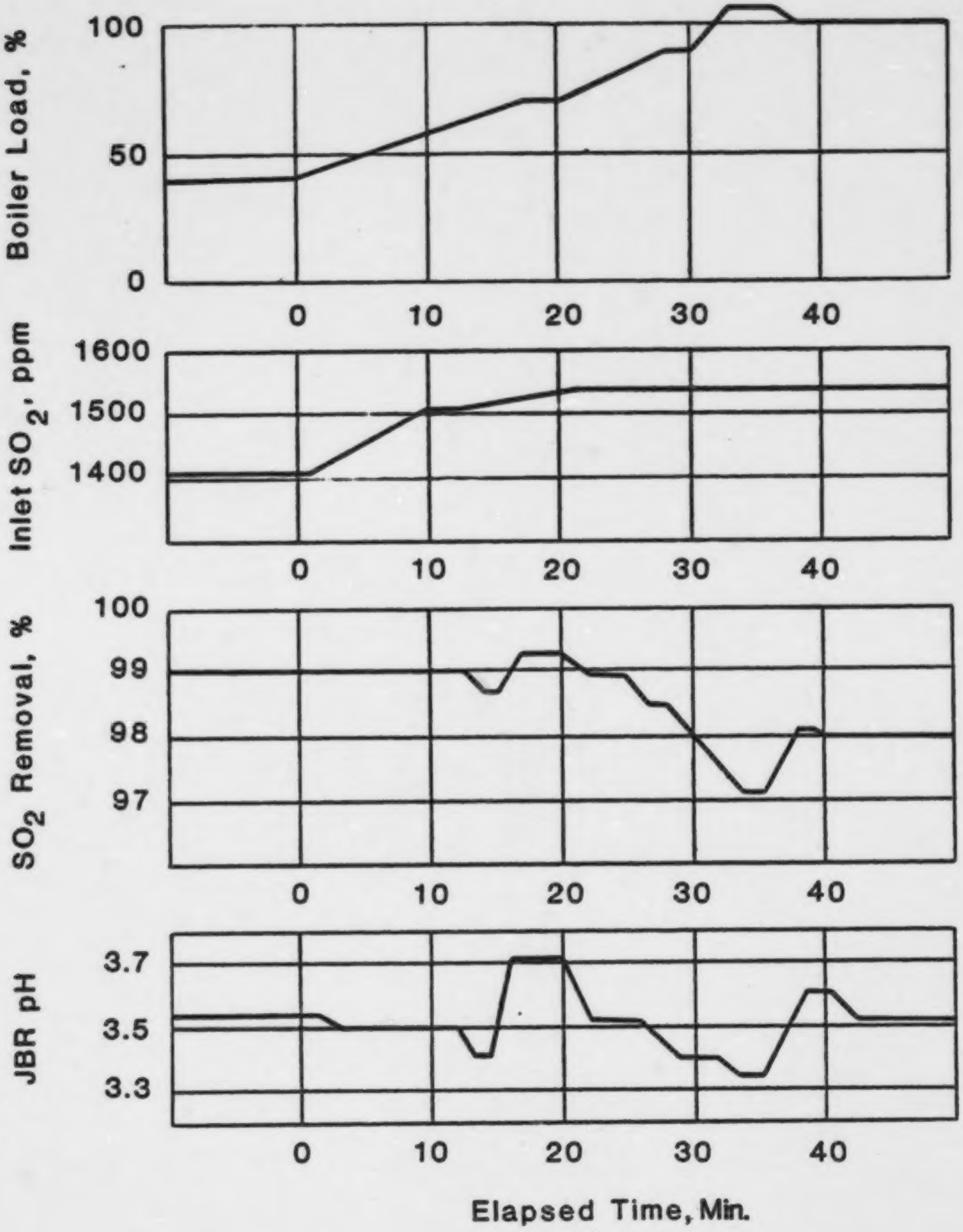

Figure 7. TYPICAL SYSTEM RESPONSE TO CHANGE IN BOILER LOAD

The pH in the JBR is controlled by the addition of limestone. The flow of limestone slurry to the JBR is controlled by the feedforward signal of the SO_2 load, compensated by the feedback signal of the JBR pH. This control loop results in only a ± 0.2% perturbation in JBR pH for maximum load changes.

POWER CONSUMPTION

The power consumption is 1,100 kW for operation at the following conditions:

Boiler Load	100% (87 MW equivalent)
Inlet SO_2	2,000 ppm
Desulfurization Efficiency	98%

Power consumption is less than 1.3% of the boiler equivalent generated MW. This consumption includes particulate removal as well as SO_2 removal. CT-121's low power consumption is due to the elimination of slurry recycle pumps characteristic of first generation limestone systems.

LIMESTONE UTILIZATION

Limestone utilization has averaged greater than 99%. Table 4 shows a typical analysis of the limestone used.

TABLE 4. TYPICAL LIMESTONE ANALYSIS

Chemical Composition	
$CaCO_3$	97.98 wt.%
CaO	54.87 wt.%
CO_2	43.80 wt.%
SO_3	0.050 wt.%
MgO	0.549 wt.%
Al_2O_3	0.059 wt.%
Fe_2O_3	0.023 wt.%
SiO_2 + Insol.	0.86 wt.%
Particle Size	
325 mesh	90% pass

BY-PRODUCT GYPSUM

A typical analysis of the gypsum by-product is shown in Table 5. The gypsum is sold to cement and wallboard manufacturers.

TABLE 5. TYPICAL GYPSUM BY-PRODUCT ANALYSIS

$CaSO_4 \cdot 2H_2O$	99.2 wt.%
SO_3	46.1 wt.%
$CaCO_3$	0.1-0.6 wt.%
Combined water	20.6 wt.%
Moisture	11-12 wt.%
pH	6.5-7.5
Crystal size (average)	49 microns

WASTEWATER

A typical analysis of the wastewater, after neutralization with limestone, is shown in Table 6. The water is discharged to the ocean.

TABLE 6. TYPICAL WASTEWATER PROPERTIES

pH	6.0
Suspended solid	0.5 ppm
COD	less than 7 ppm

OPERATING AND MAINTENANCE REQUIREMENTS

Operating and maintenance functions for the CT-121 System have been minimal and are completely absorbed by normal boiler plant operations. No scaling, plugging, or other maintenance intensive problems have been encountered. The total average personnel requirement, including maintenance, has averaged less than one-half man per shift.

RESULTS OF THE FIRST ANNUAL INSPECTION

The steam generating plant was shutdown March 1983 for the first scheduled annual inspection. The CT-121 System was also inspected at this time.

Inspection of the CT-121 System, including the JBR, revealed no scaling, corrosion, or other problems. As such, maintenance was limited to routine servicing of equipment and flushing off some soft deposits of carbon dust and gypsum crystals that had accumulated on the upper and lower decks of the JBR.

SUMMARY

The CT-121 System has met or exceeded all performance guarantees and expectations. The System has routinely achieved greater than 99% desulfurization using essentially stoichiometric amounts of limestone. System reliability has been 100%.

The CT-121 System has also proven to be an exceedingly simple process to operate and maintain. Maintenance has been minimal and limited to routine servicing of equipment. Total operating and maintenance manpower requirements have averaged less than one-half man per shift.

OPERATION EXPERIENCE WITH FGD PLANT II AT WILHELMSHAVEN POWER PLANT, WEST GERMANY

(Figure 1)

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Figure 1. Wilhelmshaven Power Plant

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ABSTRACT

The Nordwestdeutsche Kraftwerke Aktiengesellschaft (NWK) has been operating a 720 MW coal-fired power plant in Wilhelmshaven, West Germany, since 1976. In March, 1982, the second flue gas desulfurization plant, called REA 2, the German acronym for flue gas desulfurization plant 2, was put into operation after a two-year construction period. Startup and checkout proceeded without major difficulties and was completed within approximately three months. Beginning in June, 1982, NWK accepted the plant. Since then REA 2 has been in operation without significant interruptions, effectively reducing SO_2 emissions of the power plant. (Figure 2)

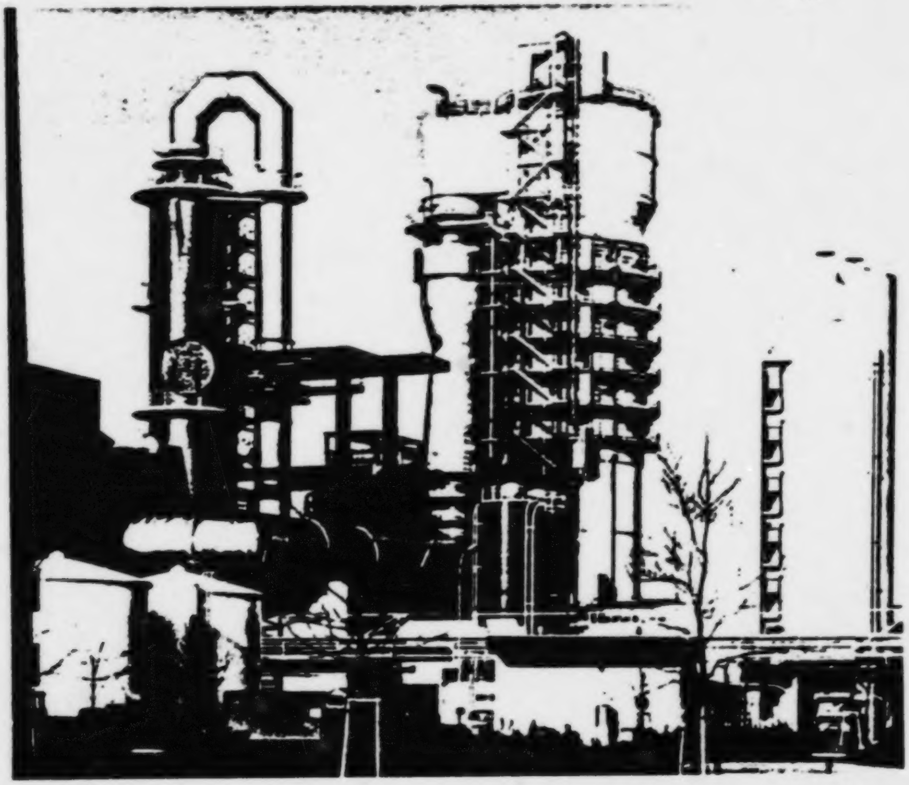


Figure 2. FGD Plants 1 & 2 at Wilhelmshaven

FGD treatment at Wilhelmshaven is accomplished by two plants. REA 1 is a co-current scrubber in operation since 1977. Together REA 1 and 2 treat about 80 percent of the flue gas generated. The FGD plants are located in the by-pass to the main flue gas duct behind the electrofilters, and are equipped with their own blowers. REA 1 has a throughput of 500,000 nm3/h (294,287 SCFM) and REA 2; 1,500,000 nm3/h (882,862 SCFM). (Figure 3)

Instrumentation for measuring dust content, SO_2 and NO_x concentrations, and exhaust temperature of the cleaned flue gas is located at the 27 m (88 ft.) platform elevation of the 275 m (902 ft.) high exhaust stack.

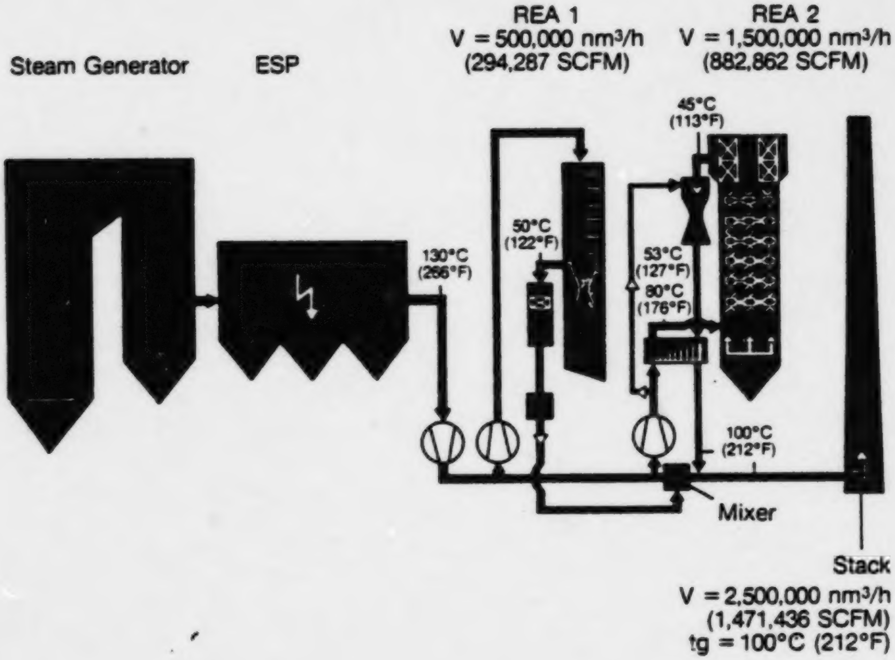
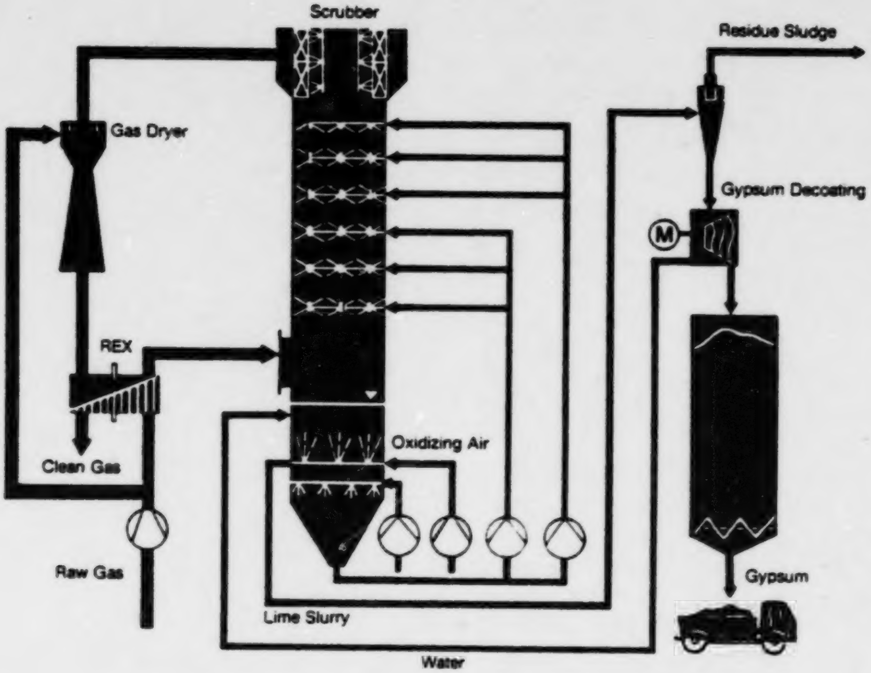


Figure 3. Present Flue Gas Flow at Wilhelmshaven

Both REA 1 and 2 use a Bischoff wet absorption process. REA 2, however, is a counter-current scrubber with regenerative re-heating of the flue gas. A portion of the flue gas generated by the power plant is injected into the REA 2 by means of an axial blower having variable vanes. From the blower, the flue gas enters the hot side of regenerative heat exchanger (REX), and from there enters from the side into the counter-current scrubber. In the scrubber, flue gas passes upward through various levels of lime slurry sprays and is cooled and desulfurized. At the top of the scrubber tower the flue gas passes through a mist separator, reverses direction, and enters a venturi mixer/preheater in which the remaining water droplets in the flue gas are evaporated by mixing hot, untreated gas with the desulfurized flue gas. From the venturi mixer, the flue gas enters the cold side of the REX and is reheated to a temperature of 100°C (212°F). From there the flue gas exits through the exhaust stack. (Figure 4)



Flow Diagram REA 2

Figure 4. Schematic of 415 MW Single-Module Bischoff System

OPERATING EXPERIENCES, PROBLEMS, AND SOLUTIONS

Operation and control functions for the two FGD plants are effected from the power plant control room. To date, no power plant interruptions have been caused by either REA 1 or REA 2. Both FGD plants can be shut down or re-started on short notice, and desulfurization occurs immediately after start-up of the blowers. In the event longer shut-down periods are anticipated, such as when a low sulfur fuel is used, for example, REA 2 can be separated from the main flue gas ducts by means of double valves, and preserved by blowing dried air into the system through both the raw and clean gas ducts.

SCRUBBER

The scrubber exhibits good operating characteristics and quickly adjusts to changes in the flue gas volume, maintaining a high degree of desulfurization. Despite a high concentration of solids in the scrubber slurry, (100 g/1), slurry liquid circulation can be readily re-started, even after being shut down over the weekend. The system does not require the use of recirculating pumps or agitators during the time it is shut down. During shut-down, about 150 tons of gypsum settle out in slurry lines, pumps and the sump. At start-up, the slurry pumps re-suspend the gypsum by backflushing for five minutes using a specially-designed system of flushing lines, after which the pumps are switched to the spray nozzles. The flushing process operates automatically.

Due to settled gypsum behind the valves, difficulties were experienced initially, in that torque overload sensors shut the automatic flushing process down when higher than permitted torques where required to open the valves. Increasing the size of the valve drive shafts and motors solved this problem.

Lime is added below the oxidation zone into the scrubber sump. The amount of lime added is controlled by means of an rpm-regulated pump. Initially, a small amount of scrubber slurry was used as the medium to add the lime. After about 2,000 hours of operation, however, the supply linesfor adding the lime clogged with gypsum. Since then, lime is added without the use of slurry fluid. The 1,000-hour operating life of the pump stators is not considered acceptable. It is hoped that this problem can be solved by installation of a bypass line directly to the scrubber sump and thus avoiding the upper rpm range of the pumps, which seems to be the cause of the problem.

During the acceptance test period, a large amount of water was found after the mist separator. The reasons were improperly installed separator segments and reintrainment of water due to open collection troughs. The former problem was corrected; the later was solved by installing covers over the collection troughs. Makeup water for the scrubber is added by using it to flush the mist separator.

To date, no scaling has been observed in the slurry systems, the scrubber, or mist separator. Nozzles have shown no wear after 4,000 hours of operation.

LIMESTONE

At the present time there are several FGD plants under construction in the Federal Republic of Germany (FRG), which, due to their proximity to limestone quarries, intend to use limestone instead of lime. In order to determine the effects in an operating plant, REA 2 was tested using limestone.

The results were as follows:

The use of limestone is possible; however, a lower degree of desulfurization is achieved than with lime. This is a result of the slower reaction rate of limestone. In order to achieve the same degree of desulfurization a considerably larger quantity of scrubbing slurry must be used. This, of course, means a higher energy requirement and greater wear of the equipment. Also, the quantity of limestone required is 1.8 times greater than that of lime due to the lower amount of CaO in limestone. However, byproduct gypsum can also be produced without difficulty.

For the Wilhelmshaven power plant, the use of limestone is not an economically viable alternative, due to the high power requirements and the greater transportation costs.

FLUE GAS DUCTING

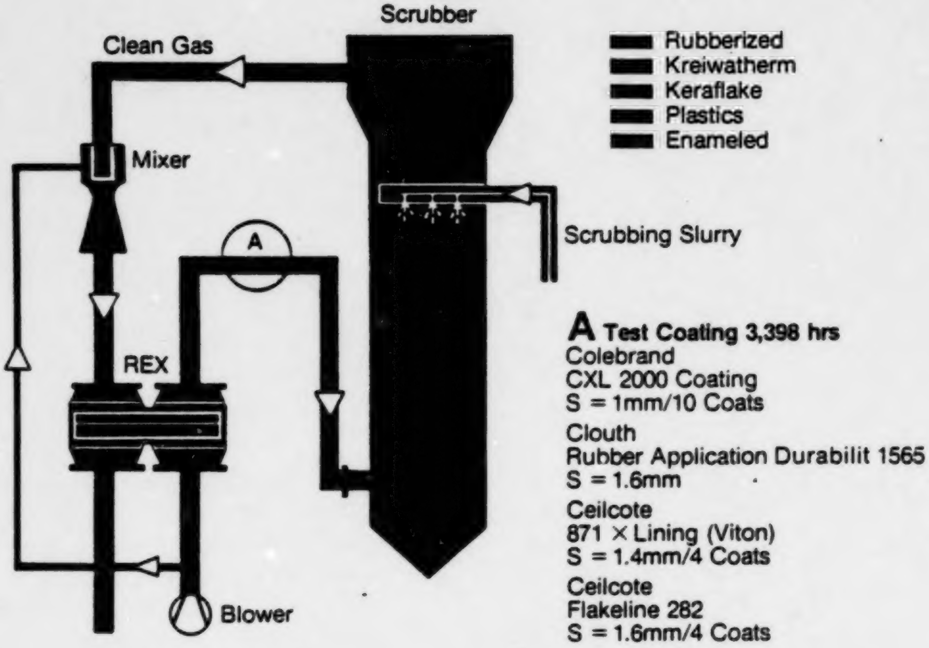
The flue gas ducting is divided into sections. Each section is free to expand or contract. Corrosion-resistant fabric expansion joints are used between the sections. During startup and testing, flue gas escaped through leaky bolt connections on the expansion joints and caused damage to the insulation. The bolted connections were improved. Since, in an FGD plant using regenerative reheat, sections of the flue gas ducting may be operated at temperatures below the dew point of the acids in gas, special precautions must be taken with respect to corrosion protection of those sections. Affected are especially the REX, the flue gas ducts between the REX and the scrubber, and the venturi mixer/dryer. Also, the flue gas ducts and especially the "floors " must be even with a slight slope and have a gutter to allow condensates, which usually have a pH of less than 1, to be drained off. The flue gas ducts between the REX and the scrubber should be sloped toward the scrubber to allow the condensate to drain into the scrubber. Other condensates flow over acid-resistant gutters to a rubber-lined collection basin from which they are pumped to the scrubber. This solution prevents an effluent problem. In case the REX must be flushed, the arrangement also allows the drainage of the flushing water to the scrubber sump.

Coating of the various components of the FGD plant presented a problem, since no experience with coating materials was available involving applications of highly concentrated sulphuric acid at temperatures greater than 80°C (176°F). Various specialty firms in this area were consulted. Only a few were willing to guarantee their coating for a minimum of two years. Several different materials were selected and applied to different sections of components of the system. After 4,000 operating hours, no final evaluation of the various coating materials is possible. Results, however, appear encouraging. It has been determined that all coatings used are satisfactory on vertical walls and on the "roof" of the flue gas ducting.

The following results are noteworthy:

The flue gas ducting should be fabricated and assembled as free of all stress as possible and should remain stress-free during operations. This is important because movements of the walls will cause the coating to crack or flake off, resulting in corrosion. The flue gas ducting should be installed such as to prevent dents, and should have sufficient slope to prevent acid puddles from forming during operations and unnecessarily stressing the coating.

The coating material must remain flexible, even after long periods of use. It is also necessary to apply the coating a sufficient number of times, as well as during warm weather. (Figure 5)



Materials

Figure 5. View of FGD Plant

INSULATION

The insulation must be thick enough and must be carefully installed. Improper insulation will result in cold spots, and thus, condensation. This in turn will result in corrosion and scaling, which can result in the discharge of solid particles from the stack. Special care must be taken in insulating support structures. It is important that in the initial request for bids, the flue gas conditions and the demands placed on the insulation be clearly spelled out.

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REGENERATIVE HEAT EXCHANGER (REX)

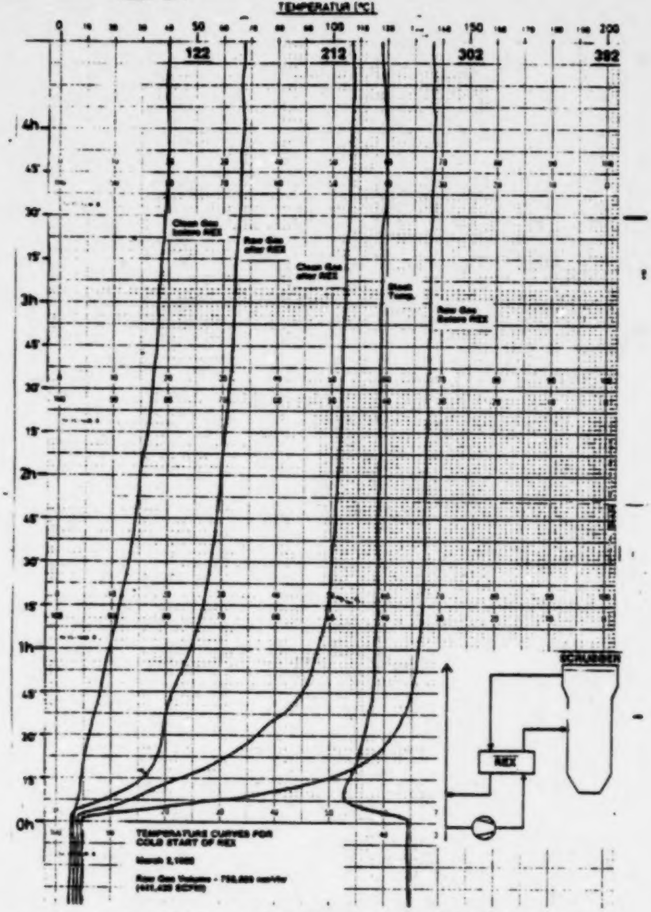
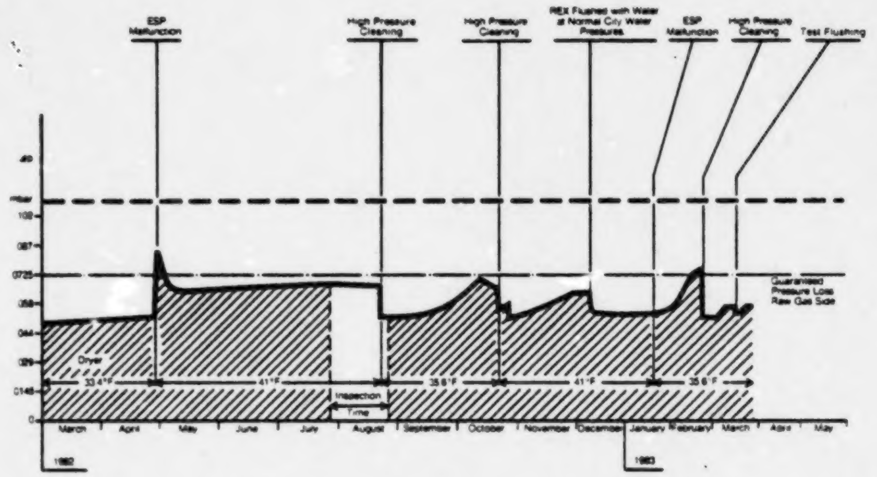


Figure 6. Temperature Profile of REX for Cold Start

The concept of reheating the desulfurized flue gas has been viable todate. Startup of the REX, from a cold condition, has not presented any problems. Gas discharge temperatures, measured in the stack, always exceed 100°C (212°F), even during periods of partial loads. The heat-storage chambers are made of plastic material on the cold side and of enamel-coated steel on the hot side. On the cold side, deposits form, caused by water droplets entrained in the clean gas. These evaporate at the inlet of the REX. On the hot side, deposits form due to entrained dust particles containing acidic deposits in the raw gas. (Figure 6)

In case the electrostatic precipitator does not function properly, the resulting increase in ash content of the raw gas quickly increases the pressure drop across the REX.

The hot surfaces are cleaned by means of pressurized air, which is blown over the surfaces by special nozzles. This is performed after every 12 hours of operation. During normal operations, i.e. without malfunctions in the ESP or the scrubber, this cleaning interval has been found to be sufficient. Should larger deposits form more quickly due to malfunctions in either the EXP or the scrubber, cleaning by means of high pressure water is required. The same nozzles used for the high pressure air are also used for the high pressure water spray. For this purpose, they are switched to a high pressure water pump. On the hot side, the water pressure before the nozzles is 80 bar (1160 psi) and 40 bar (580 psi) on the cold side. The REX continues to turn during the cleaning process with a reduced velocity. The amount of water used is held to a minimum. This water is added to the scrubber as make-up water. The cleaning method has been used during operation of the REA 2, as well as during shutdowns.



Pressure Loss of REX During Operating Period

Figure 7. Pressure Drop Diagram For REX

Measurements to determine pressure losses across the REX from start-up until the 29th of March, 1983, show that the pressure loss increase has been smaller than anticipated. The deposits on the REX have been maintained in acceptable limits during the first 4,000 hours of operation by cleaning regularly with high pressure air or water. The installed reserve capabilities of the blower, layed out for a pressure drop increase up from 10 mbar (0.145 psi) to 13 mbar (0.216 psi), has been sufficient. (Figure 7)

Tests will soon be conducted to determine whether deposits on the hot side can be effectively removed using steam instead of water. To date, no noticeable corrosion has been determined in the REX.

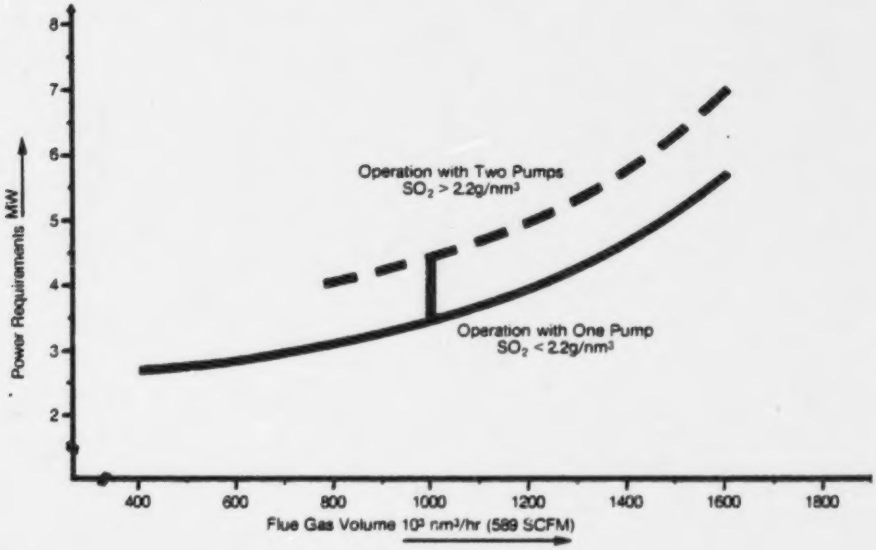
No definite statement can be made regarding the lifetime of the heat-storage chambers after this short period of operations.

PRESSURE LOSSES

Measurements of pressure losses across various components of the REA 2 system at start-up showed that the pressure drops were below those given by the various suppliers. Pressure losses across the scrubber and the flue gas ducts do not change with time.

ENERGY REQUIREMENTS

Testshave shown that in those cases when the SO_2 content of the flue gas is in the lower end of the operational spectrum (SO_2 less than 2 g/mn^3); (0.8 gr/scf) or flue gas volume less than 1,000,000 nm^3/h (588,574 scfm), the desulfurization efficiency does not increase significantly if two slurry pumps are operated. Using only one slurry pump, the desulfurization efficiency will only decrease by 2 percent; however the power saved is 1.2 MW. (Figure 8)



Energy Requirements of REA 2

Figure 8. Energy Requirements for REA 2

INSTRUMENTATION AND CONTROLS

Control

For controlling the REA 2, it was decided for the first time in this power plant to use a programmable controller. The controller used is of the Siemens type 5 and is used in conjunction with a more conventional control system, ISKAMATIC B. The S 5 system is used to control the functional groups, including the safety interlocking systems. The great advantage offered by the programmable S 5 system is the ease with which the automatic controls can be adjusted to the operational process. This was shown to be of great advantage during the start-up phase of the operation.

The following group controls are used:

- flue gas blowers with oil supply and valves
- slurry pumps with flushing program
- flushing of mist separator
- lime addition

Modifications are presently under way to upgrade the automatic control system to also handle clarifying and gypsum dewatering. It is intended to control the entire FGD plant from the power plant control room and includes starting the plant and shutting it down.

Performing the various controlling functions is accomplished with the process control system TELEPERM C. Control is accomplished using the following control loops:

- raw gas volume
- drying of the clean gas
- controlling amount of lime added
- solids removal, slurry discharge
- control of various fluid levels

Since the REA 2 is a counter-current scrubber, it is, theoretically at least, possible to adjust the pH value at the slurry discharge to such a value as to optimize the oxidation of the sulphite to gypsum, without adversely affecting the desulfurization.

This results in the following very important advantages. First of all, the lime usage is stoichiometric, and readily adjustable as operating parameters change. Second, it is possible to integrate the oxidation stage in the scrubbing tower. This, of course, does increase the demands placed on the instrumentation and control system, since, unlike the co-current system, the scrubbing fluid is practically unbuffered.

The pH control loop was modified during start-up. Instead of continuously changing lime dosage as a function of pH, lime dosage is now adjusted in discreet steps. However, more recent tests indicated that, by also measuring density, a more accurate mass flow rate of the lime to be added is now possible. Originally, a rough dosage and a fine dosage were used for the injection of lime required. The fine dosage was found to be not required and was disassembled.

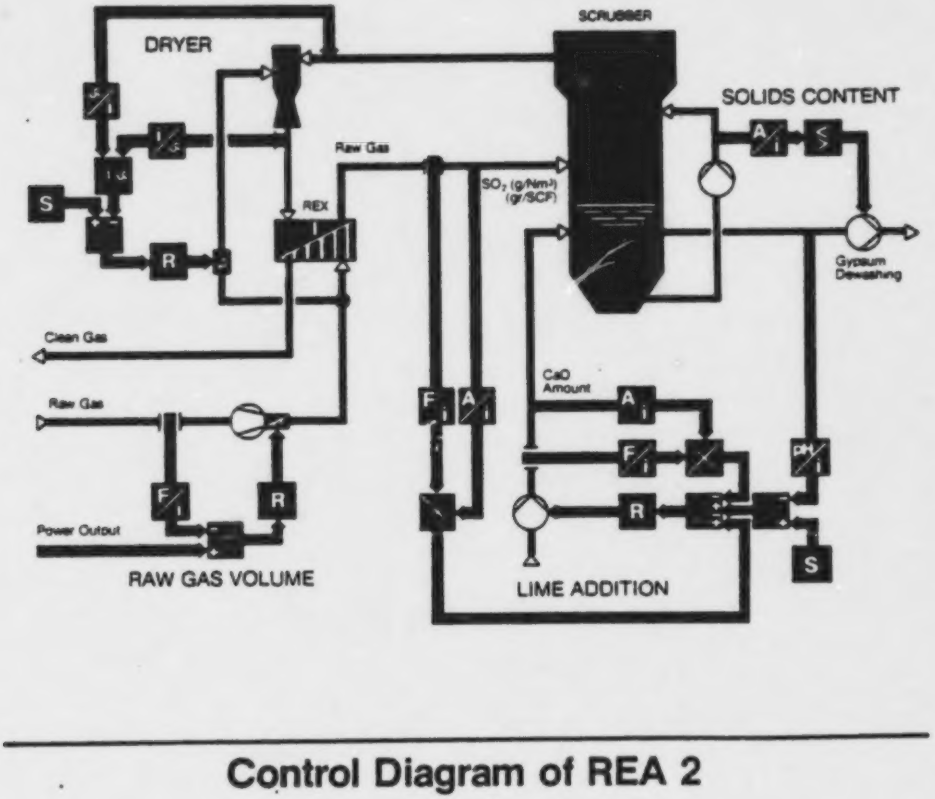


Figure 9. Control Diagram

INSTRUMENTATION

In order to monitor and automate the operation of the FGD plant, as well as to optimize the dosage of additives and the desulfurization, reliable and maintenance-free instruments must be used. This is especially true for determining the amounts of additives supplied to the system by measuring the solids content in the slurry and the amount of lime added, the pH value of the wash fluid and the SO_2 content in the raw and clean gas.

The following systems were selected:

- solids and density measurements:	radiometric analysis using the absorption principle
- SO_2 concentration measurement	spectral photometer using the absorption principle
- pH value measurement	electrometric analysis, potentiometer
- flow measurements	induction measurement
- scrubber fluid measurement	ultrasonic measurement
- flue gas volume measurement	pito tube (annular tube)

Most of these instruments have proven to be very reliable. Determination of SO_2 concentration using the absorption principle causes some problems, however. (Figure 9)

EFFICIENCY OF SO_2 REMOVAL

The degree of SO_2 removal in the scrubber is influenced by various factors. These are: the relationship between scrubbing fluid volume and gas volume;gas velocity; pH value; and SO_2 concentration of the raw gas. During acceptance tests, it was found that 95 percent of the SO_2 in the gas was scrubbed out. Under operational conditions, with SO_2 concentrations of between 1 g/nm3 and 3 g/nm3 (0.4 gr/scf and 1. gr/scf), 91 percent to 95 percent of the SO_2 was removed. Because the regenerative gas reheater was selected in the REA 2, the overall SO_2 removal efficiency is lower than in the scrubber, due to the fact that raw gas is added to the clean gas after the mist separator to evaporate any water droplets still in the gas, and also due to the fact that some raw gas leakage occurs in the regenerative gas reheater.

Drying of the clean gas, using raw gas, was selected to prevent deposits from forming on the regenerative gas preheater.

To evaporate the rest moisture in the clean gas, the clean gas temperature must be raised. The rate of evaporation, due to the increase in temperature, is determined by the available transit time of the gas through the dryer. It must also be considered that the salt content of the evaporated moisture will increase the salt concentration, resulting in decrease of the water vapor pressure and thus increasing the dew point.

The amount of increase in the dew point depends on the type of salt desolved. Table salt, magnesium chloride and calcium chloride result in dew point increases of about $5^{\circ}C$ ($9^{\circ}F$), $30^{\circ}C$ ($54^{\circ}F$), and $45^{\circ}C$ ($81^{\circ}F$) respectively, when present in saturated solutions, Drying, by using raw gas, is thus only viable when readily soluble solutions such as magnesium of calcium chloride are not present. At Wilhelmshaven, where the REA 2 is operated with sea water, this does not appear to be a problem.

A delta T of $8^{\circ}C$ ($14.4^{\circ}F$) is achieved in the gas preheater, resulting in a decrease in the SO_2 removal efficiency of about 10 percent. This disadvantage of clean gas drying can only be avoided if a highly efficient mist separator can be used, decreasing the need for such a dryer by removing a sufficient amount of the residual moisture and salt.

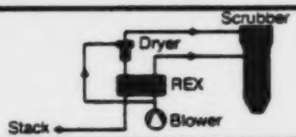
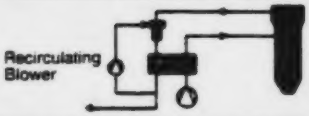
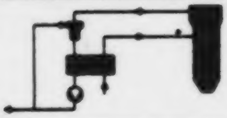
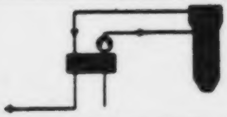
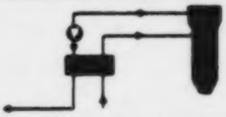
	Δ T (Dryer)	Decrease of Scrubber Desulphurization Efficiency
	Power Plant Wilhelmshaven	
	8°C (46.4°F) 2°C (35.6°F) 0°C (32°F)	12.5% 5 % 2.5%
	Drying with Recirculating Blower	
	2-5°C (35.6-41°F)	2.5%
	Blower Using Hot Clean Gas	
	2-5°C (35.6-41°F)	2.5%
	Blower Using Cold Raw Gas	
	-	1%
	Blower Using Cold Clean Gas	
	~3°C (37.4°F) Due to Blower	1%

Figure 10. Total Desulfurization Efficiency of REA 2 with REX

By decreasing the delta T from $8^{\circ}C$ ($14.4^{\circ}F$) to $5^{\circ}C$ ($9^{\circ}F$), no effects on the REX have been determined. During the winter of 1982, a delta T of $2^{\circ}C$ ($3.6^{\circ}F$) was measured across the gas dryer. In this case, deposits began to build up on the surfaces of the cold side of the REX. These deposits could not be removed by means of high pressure air; they were, however, removed with high pressure water.

Thus, a $2^{\circ}C$ ($3.6^{\circ}F$) delta T in the dryer is possible if cleaning of the REA with high pressure water is acceptable. Increased mist separator efficiencies would increase operating times between flushing considerably.

Without the use of the venturi dryer, no noticeable difference in the increase in pressure drop across the REX resulted, as compared to operation with the dryer. These tests encompassed a time period of two months of effective operation. In this case, only normal air cleaning at 12-hour intervals was employed. Tests are presently continuing.

It is not possible, after the relatively short time since this system went into operation, to determine if flushing with high pressure water will cause damage to the REX due to increased corrosion or loosening of the regenerative heater packages.

By reducing the delta T from 8°C (14.4°F) to 5°C (9°F), SO_2 removal efficiencies increase from 82 percent to 89 percent. By not using the dryer, the efficiency could be raised to 92 percent. If this is not possible the gas would have to be dried using the blower on the clean gas side. This would not affect the efficiency of the scrubber.

In such a case, gas leakage in the REX does not decrease the SO_2 removal efficiency of the system significantly. The resulting increase in the clean gas temperature caused by the blower minimizes the need for a gas dryer. Such a solution is presently being contemplated. (Figure 10)

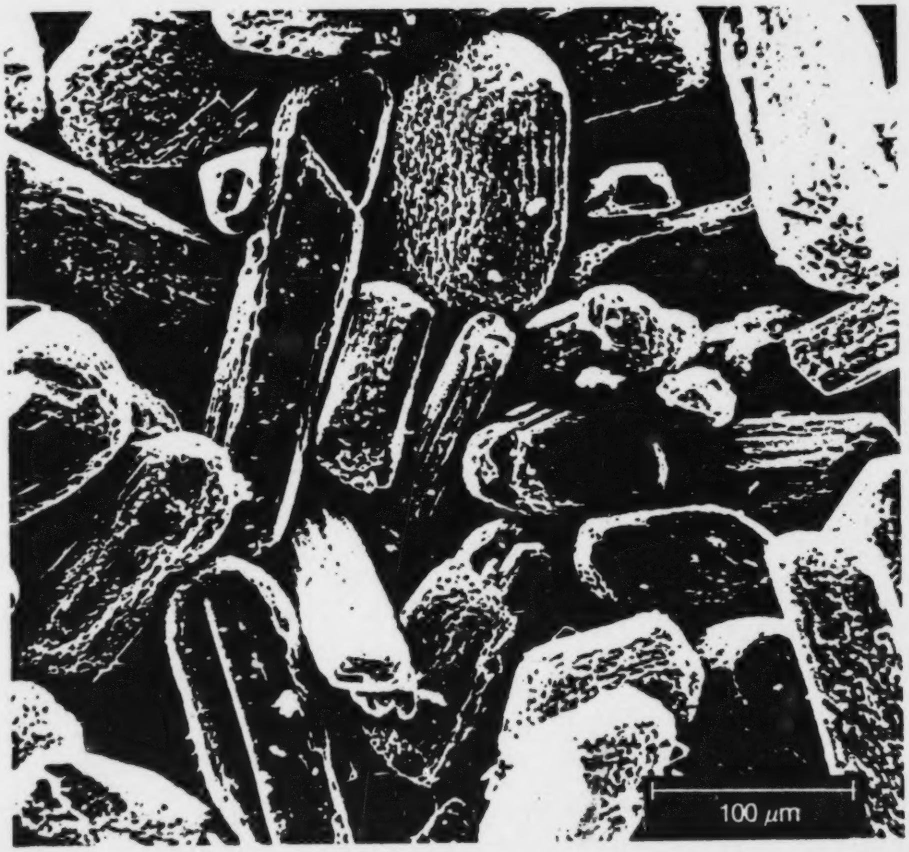


Figure 11. Magnified Photograph of the Crystals

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GYPSUM PREPARATION

The byproduct generated by the Bischoff FGD plant is high quality gypsum containing a high proportion of large gypsum crystals. Because of this, NWK and a wallboard manufacturer have signed a contract to utilize the FGD byproduct gypsum for wallboard production. The contract specifies the quality of the gypsum. To meet these specifications, the byproduct gypsum must be washed and dewatered. (Figure 11)

In purchasing the equipment necessary to wash and dewater the gypsum, cost was an important factor, since the profit resulting from the sale of the gypsum should result in a reasonable payback period. The equipment presently being installed or planned is complicated and expensive. In general, such a system consists of various intermittently-operating centrifuges, or continously-operating vacuum filters with an additional system to transport the gypsum to a silo or other storage facility without additional handling.

To find a suitable system, extensive tests were conducted with the following systems:

- vacuum-band filter
- vacuum-drum filter
- two-stage centrifuge
- four-stage centrifuge
- screw centrifuge

All of these systems are, under certain conditions, suitable for use in washing and dewatering byproduct gypsum. The following residual moisture contents are measured.

vacuum-filter	8 - 10 %	(without free water)
centrifuge	6 - 8%	(without free water)

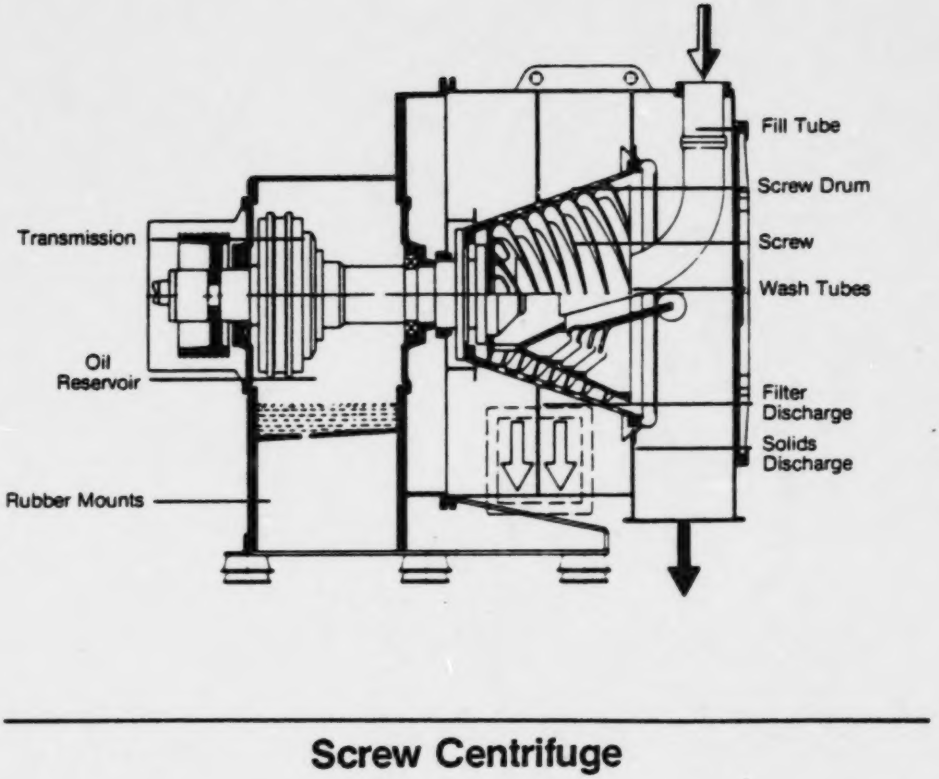

Figure 12. Screw Centrifuge

At Wilhelmshaven the screw centrifuge was selected due to the fact that it was less expensive, required less space and energy, could be automated without much difficulty, and finally, could be expected to have low-wear characteristics. (Figure 12)

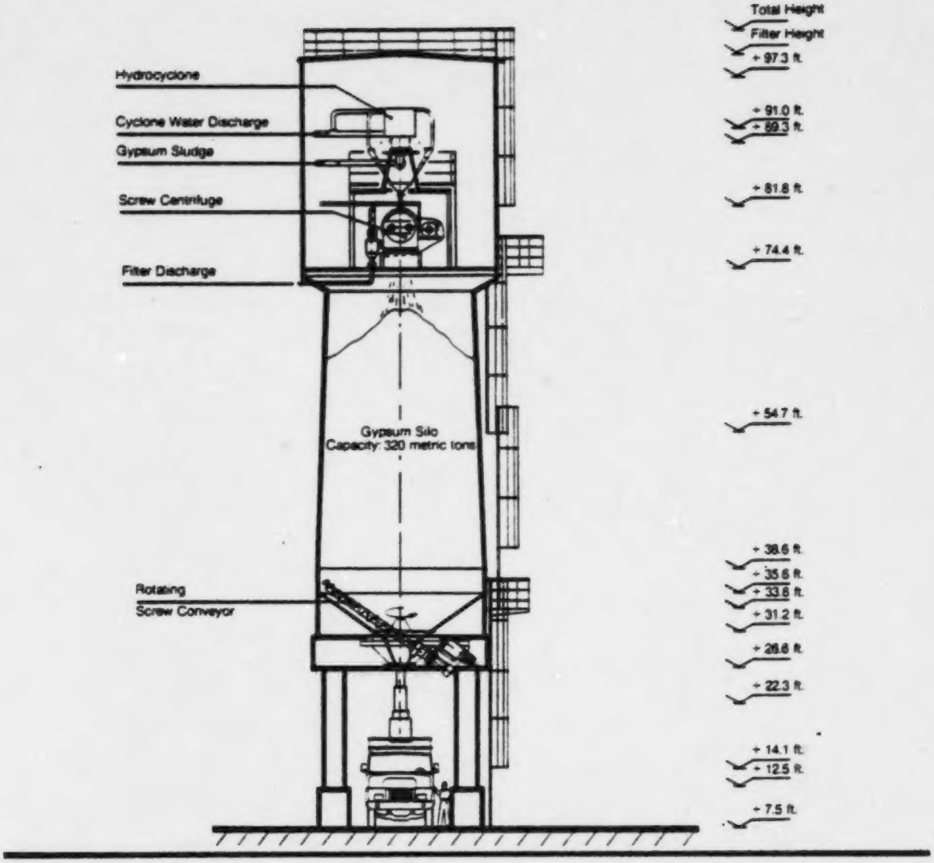

Figure 13. Gypsum Silo with Centrifuge

The centrifuge is installed over a silo, which has a capacity of 2½ days of gypsum production. Centrifuge, silo, and loading facility are one structure. Loading is accomplished by means of a screw conveyor which is activated by the truck driver. Before the gypsum slurry reaches the screw centrifuge, fine particles are removed by a hydrocyclone. A portion of the separated material flows to a thickener. The main stream is returned to the scrubber. The gypsum preparation plant is started up, or shut down, depending on the measured solid content in the scrubber slurry. Boundary values are 60 to 120 g/l. (Figure 13)

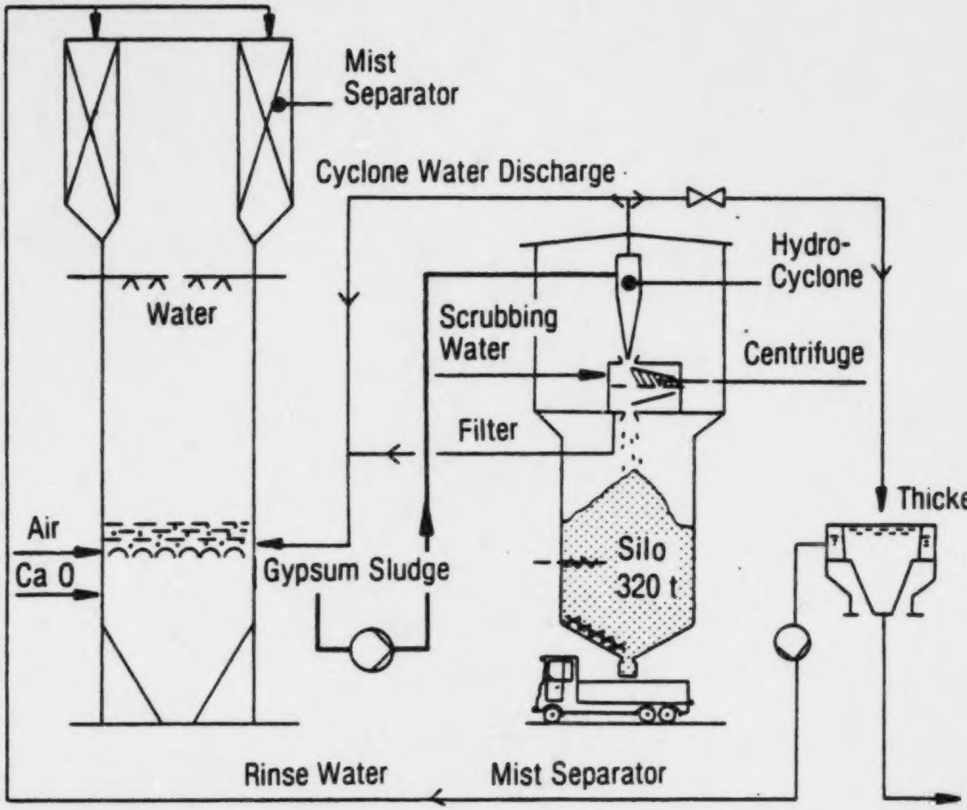

Figure 14. Schematic of Gypsum Preparation Plant

The gypsum generated in 24 hours of desulfurization can be withdrawn from the scrubber within eight hours. Gypsum is withdrawn when the solids content in the slurry reaches 123 g/l. Withdrawal ends when the solids content reaches 60 g/l. To insure that only large gypsum crystals are removed, the centrifuge filters out undersized crystals and returns them to the scrubber. In the time required for the solids content to again reach 120 g/l, the gypsum crystals have ample time to grow. Thus, an adequate supply of large crystals is always assured. (Figure 14)

The cost for the REA 2, including the gypsum preparation plant, was about $25,000,000.00. The various components, such as washer, REX, blowers, gas ducting, instrumentation and controls, as well as installation, were selected and ordered by the power plant. Engineering of the plant was accomplished by an engineering company.

SUMMARY

The availability of the REA 2 meets our expectations. Scaling the REA 2 from REA 1, which is 1/3 as large, presented no problems. As a result, a FGD plant capable of serving an 800 MW power block was developed.

Regenerative reheat has been proven to be possible, and it saves energy. Definitive conclusions regarding corrosion can not as yet be made.

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THE SULF-X PROCESS

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ABSTRACT

The purpose of this paper is to report the status of development of SULF-X technology for flue gas desulfurization and to review the design and applicability of the Process in boiler service firing diverse coal fuels. Details are given of process chemistry, design and operation, economics and past and future system demonstration programs. The flexibility and attractiveness of the technology is shown in its use either for SO_2-removal-only or for simultaneous SO_2/NO_x removal.

INTRODUCTION

The SULF-X Process is a wet absorption process that utilizes a slurry of regenerated ferrous sulfide solids to achieve removal of 90 to 99% of sulfur dioxide from boiler flue gases by wet scrubbing. It is technically feasible for use with all fossil-fuel types. A coal-fired calciner regenerates both spent and oxidized forms of the ferrous sulfide reagent, thereby converting collected sulfur dioxide into salable elemental sulfur while minimizing formation of non-regenerable waste solids and liquids requiring disposal.

Current technical evaluation and economic analysis of the SULF-X technology by the Electric Power Research Institute shows that it is substantially less costly than the familiar Wellman-Lord regenerative flue gas desulfurization (FGD) process.(1) The study also indicates that it may be more attractive than limestone-based throwaway solid-waste type systems, particularly at sites where waste management options are limited. The process is primarily applicable in high sulfur coal-burning regions where marketing of a commerical grade of by-product elemental sulfur is feasible and where industrial by-product iron compounds such as pyrites (ferrous disulfide) are available for use as process reagent makeup supply. With the benefit of revenues from sulfur by-product sales of $150 or more per short ton and availability of inexpensive pyrites or copperas (ferrous sulfate heptahydrate), SULF-X is highly competitive with limestone-consuming FGD systems in common use. Pyrites may be derived from coal cleaning or from mining of non-ferrous metals. Copperas is commonly available from industrial waste processing including waste pickling liquor neutralization.

PROCESS CHEMISTRY

The reaction steps in the SULF-X Process are schematically illustrated in the simplified block diagram (Figure 1).

WET SCRUBBING

SO_2 Removal

An aqueous slurry containing a complex mixture of iron/sulfur compounds, including finely divided ferrous sulfide (FeS) reagent in stoichiometric excess, is used to absorb sulfur dioxide in a wet scrubber. Although the solubility of FeS in water is only 6 mg/liter (at 68°F), it reacts with absorbed SO_2 and oxygen converting a portion of the SO_2 to elemental sulfur while being sulfidated and, in part, degraded to soluble ferrous sulfate. Additionally, while the individual absorption reactions are numerous and complex, the overall chemical reaction in absorption of SO_2 may be expressed as follows:

$$(c)\ FeS + SO_2 + O_2 \rightarrow (c-1)\ FeS_{(m)} + Fe^{++} + SO_4^{=} + (1 - \frac{n}{10})\ S^0$$

The quantity "n" can range from 1 to 10 depending on process conditions. The quantity "c" is greater than 1 but less than 2, its magnitude varying depending on equilibrium among the diverse chemical reactions taking place. In a typical design, rate of FeS feed to the absorption reaction is based on the assumption that two moles are required per mole of SO_2 absorbed (c = 2). The quantity "m", which ranges between 1 and 2, is a measure of the degree of sulfidation of the regeneated FeS and reflects the spectrum of possible reaction products between FeS (fresh/regenerated) and FeS_2 (fully spent reagent). The SULF-X Process is typically designed and operated to limit elemental sulfur formation in the wet-scrubbing step (to maximize "n") so that difficult to separate finely suspended sulfur will be yielded in the subsequent thermal regeneration step where it is more easily isolated and recovered.

Intermediate absorption reaction products include HS^-, $SO_3^=$, HSO_3^- and $S_2O_3^=$ (thiosulfate). Secondary reactions not reflected in the overall absorption reaction equation include oxidation of ferrous ions to ferric and oxidation of bisulfite and sulfite ions to sulfate.

NO_x Removal

The equation for the SO_2-removal reaction above illustrates the oxygen absorption capability of the process. Through retrofit modification or by provision in the initial installation, the SULF-X Process may be designed to use this chemical reducing (deoxidizing) characteristic of its process slurry to achieve simultaneous removal of NO_x from boiler flue gas. Catalysis necessary to accomplish this is provided by operating the system so as to maintain approximately 40 grams per liter ferrous ion (Fe^{++}) concentration in the continuously recirculating scrubbing slurry. The overall chemical reaction in this simultaneous absorption of SO_2 and NO_x may be expressed by the following simplified single equation:

$$(c)\ FeS + SO_2 + NO + \tfrac{1}{2}\ O_2 \xrightarrow{Fe^{++}} (c-1)\ FeS_{(m)} + Fe^{++} + SO_4^{=} + (1 - \frac{n}{10})\ S^0 + \tfrac{1}{2}\ N_2\uparrow$$

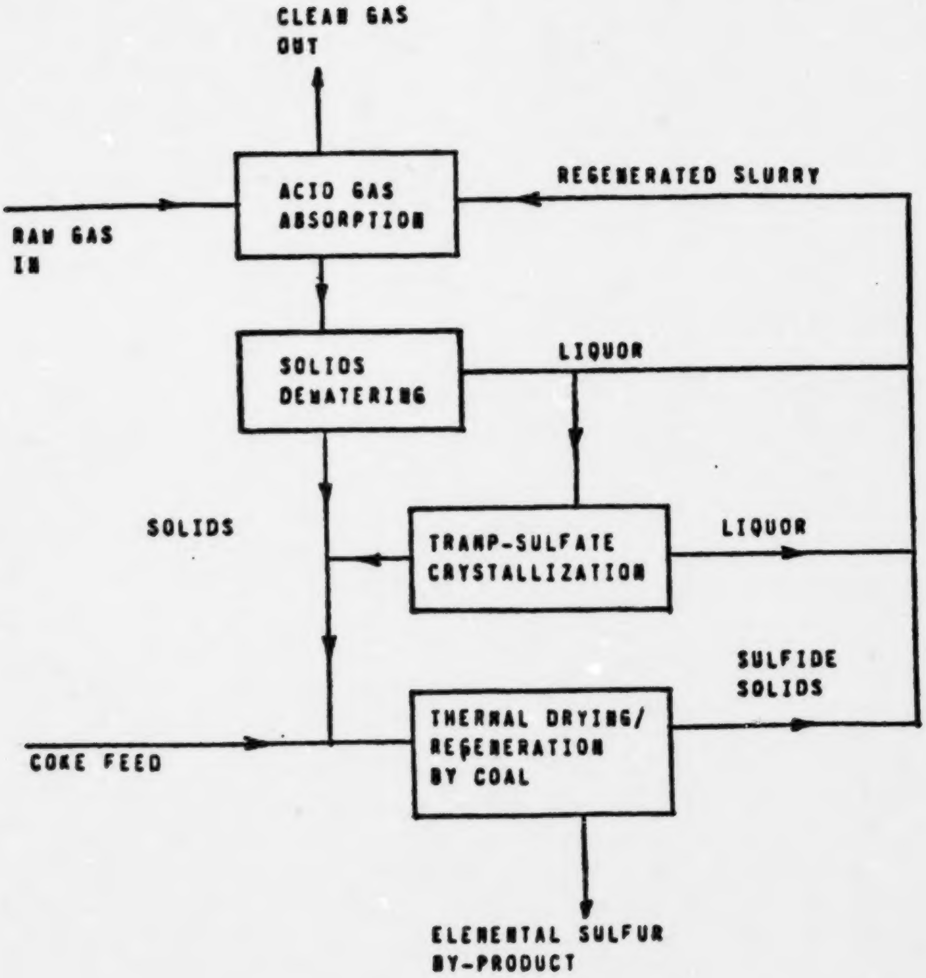

Figure 1. SULF-X Process Flow Diagram

Additional intermediate reaction products include $Fe(NO)^{++}$ formed in initial capture of NO. To remove most of the NO_x, substantially greater reaction time is needed in the absorber than is required in collection of SO_2 alone. Thus, for simultaneous SO_2/NO_x removal, the installation must provide, either through retrofit modification or by initial design, additional absorber capacity to increase the gas residence time.

Note that the absorption-reduction environment in the process slurry chemically reduces the collected NO_x to innocuous elemental nitrogen. By contrast, a simultaneous SO_2/NO_x removal process based on absorption-oxidation chemistry that utilizes an oxidizing agent, converts collected NO_x to potentially water-polluting oxidized forms such as nitrites (NO_2^-) and nitrates (NO_3^-).

Absorption-reduction can be distinguished from absorption-oxidation by a comparison among familiar types of calcium-alkali scrubbing systems. An absorption-reduction environment exists in some of the common scrubbing systems such as those using magnesia-buffered lime and sodium-liquor scrubbing (dual-alkali FGD) that operate in an unsaturated-$CaSO_4$-mode with a significant ionic concentration of sulfite/bisulfite ($SO_3^=/HSO_3^-$), an oxygen scavenger. On the other hand, lime/limestone FGD systems, as most commonly applied, function without appreciable dissolved concentrations of such chemical deoxidants. Their scrubbing slurries typically impose an absorption-oxidation environment due to the presence of dissolved oxygen and are incapable of chemical reduction.

SULFIDE/SULFATE REGENERATION

Suspended solids in the slurry bleed-off from the acid gas absorption step (Figure 1) are dewatered prior to roasting in an indirect-fired calciner. Most of the process liquid from bleed dewatering, principally containing dissolved sodium sulfate (Na_2SO_4) and ferrous sulfate ($FeSO_4$), is used as a quench to cool and slurry the hot solids leaving the calciner. The balance of the liquid is diverted to a crystallizer system to recover sodium sulfate solids. The spent scrubber solids combined with the crystallizer solids and a proportion of coke constitute the feed to the calcining step. In the calciner at 1200 to 1400°F, sulfide is regenerated and sulfur by-product is formed in accordance with the equations as follows:

$$FeS_{1+n/10} \xrightarrow{\Delta} FeS + \frac{n}{10} S^{o}\uparrow$$

$$Na_2SO_4 + 2\,C \xrightarrow{\Delta} Na_2S + 2\,CO_2\uparrow$$

To avoid oxidation of any of the sulfur yield, excess coke is supplied to the calciner to provide a protective reducing atmosphere containing carbon monoxide (CO). During the quenching step noted above, the sodium sulfide in the calcine forms FeS precipitate:

$$Na_2S + FeSO_4 + H_2O \rightarrow FeS\downarrow + Na_2SO_4 + H_2O$$

THROWAWAY WASTE GENERATED

The SULF-X Process produces a minimum of non-regenerable wastes and thereby minimizes the environmental impact of acid gas emission control facilities.

Solid Waste

Oxidation of scrubbing slurry results primarily in the conversion of sulfides to sulfate, which is regenerated to sulfide by calcination. Thus, the quantity of throwaway solid waste is negligible in comparison with the amount of ash and other solid wastes generated by coal-fired boiler operations. A comparatively small proportion of ferrous iron, estimated to be equivalent to less than 5% of the stoichiometric SO_2 removal rate in pulverized coal-fired boiler applications, is oxidized to ferric oxide (Fe_2O_3) in the scrubbing step. Ferric oxide is a tramp solid material that may be discarded from the system through disposal of a small amount of dewatered scrubber-bleed solids. However, the reducing conditions in the calcination step convert a large portion of the ferric oxide solids to reactive ferrous compounds, thereby decreasing the net amount of iron oxide waste production.

Liquid Waste

Chloride Management--

As in other wet scrubbing processes, hydrogen chloride contained in the inlet flue gas is efficiently absorbed and must be purged in a liquid effluent.(2) To prevent loss of iron salts in this purge stream and to minimize complexity of wastewater management, the absorbed chloride may be segregated in an isolated pre-scrubbing loop as is done in application of other regenerative FGD processes.

Other Waste-ion Formation--

Chloride is the only non-regenerable dissolved solid that tends to accumulate in the system. Thiosulfate ($S_2O_3^=$), a common tramp compound that must be purged in sodium-base regenerative FGD systems, is both formed and reconverted to bisulfite (HSO_3^-) in the SULF-X scrubbing step. Thus, purging of thiosulfate from the system is not required. Moreover, unlike sodium sulfite based absorption-reduction processes utilized for simultaneous SO_2/NO_x removal, the SULF-X Process does not form detectable quantities of dithionate, imidodisulfonate or other nitrogen/sulfur tramp ion complexes.

SYSTEM DESIGN AND OPERATION

The general flow diagram for a commercial SULF-X Process installation in coal-fired service is illustrated in Figure 2.

OVERVIEW OF DESIGN BASIS

Gas Absorption

As in current typical application of other flue gas desulfurization processes, an electrostatic precipitator (or fabric filter), not shown in Figure 2, is used to control stack fly-ash emission and to limit the amount of solid particulates entering and accumulating in the wet scrubbing operation. In addition, a prescrubber loop as shown is used upstream when required for absorption of other tramp materials, including HCl. This limits their impact on the operation of the process and the control of process chemical losses (through closed-loop recycle of SULF-X liquid and solid flows). Design of large-scale SULF-X absorbers is tied closely to the prior development and continued test/demonstration activity in the simultaneous removal of SO_2 and NO_x. To ensure mass transfer capacity sufficient to meet potential NO_x control objectives, the packed

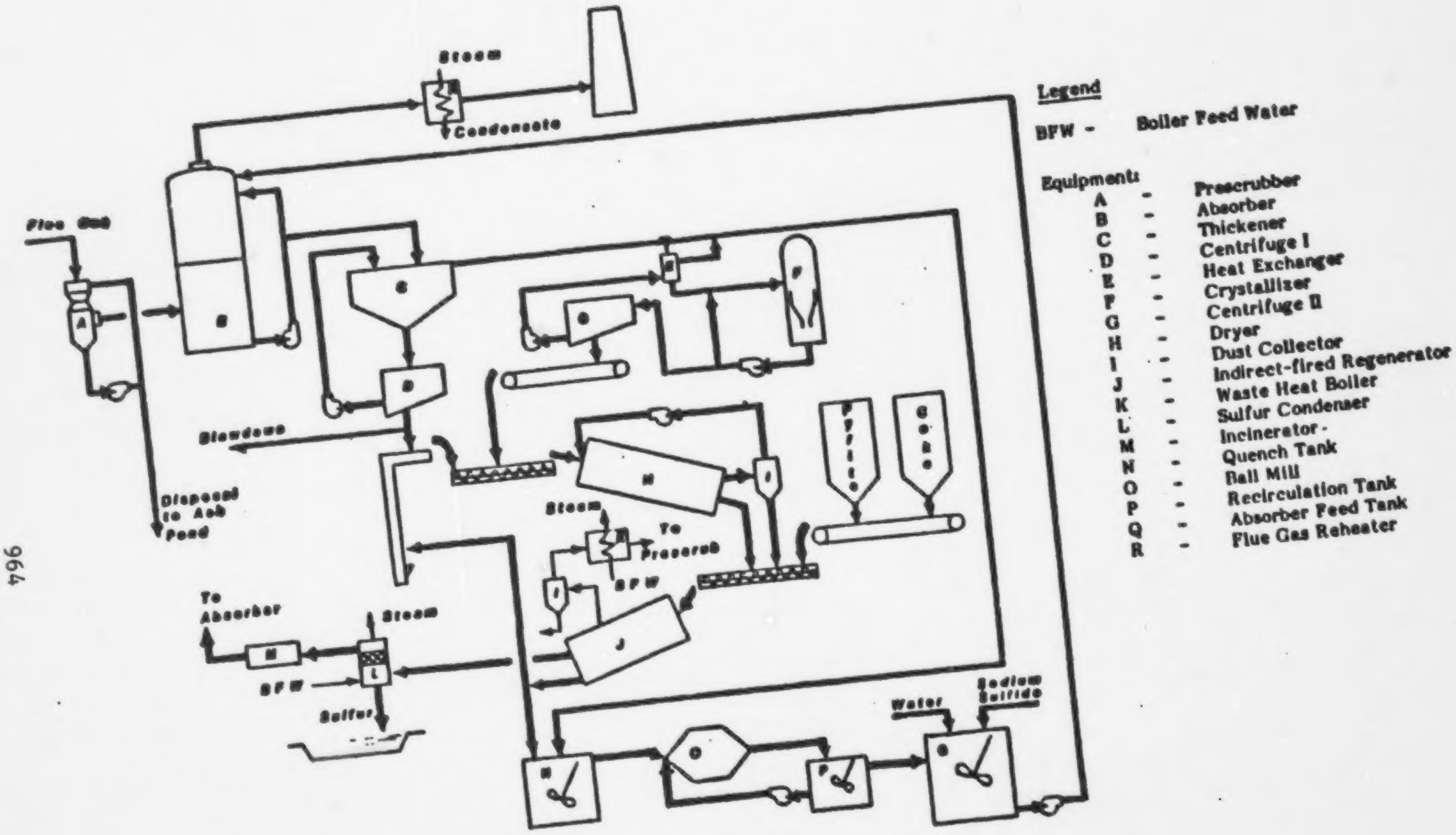

Figure 2. SULF-X Process Flow Diagram

absorber design utilizes fixed cross-fluted Munters packing material in counter-flow scrubbing operation. In the design of the packed absorber in SO_2-only service, a recirculating slurry to gas flow ratio of approximately 60 U.S. gallons per thousand actual cubic feet is used with an absorber gas-pressure-drop of 8 in. w.g. at a design superficial gas velocity of 10 feet per second. In such applications concerned solely with SO_2 removal in the 85-95% efficiency range, a spray-tower-type absorber, operated at a superficial gas velocity comparable to designs used in lime/limestone slurry scrubbing systems, may be found to be preferable. Due to provisions in the process for regeneration of sodium sulfate, the principal product of parasitic oxidation in the absorber, the SULF-X absorption-reduction system can, uniquely, be applied to emission sources such as bituminous-coal-fired stoker boilers that may contain oxygen concentration in the raw flue gas as high as 14%. Thus, with this capability for processing flue gases with a very high O_2/SO_2 ratio, the SULF-X Process can be applied to pulverized coal boilers fueled with coal of all ranks and with either high or low sulfur content.

Sodium Sulfate Cystallization

An atmospheric pressure crystallizer column is used to preferentially crystallize the sodium sulfate make for regeneration to sulfide in the calcining operation. Reasons supporting this method for recovering and reusing sulfate waste product are as follows:

- Chemical reduction of sulfate is thereby accomplished without liberation of SO_2 that accompanies thermal processing of alternative sulfate solids forms such as $FeSO_4$.
- Iron compounds that are present catalyze the reduction of Na_2SO_4.
- The minimum temperature for reduction does not exceed the design temperature for simultaneous calcination of the spent FeS solids ($FeS_{1 + n/10}$ forms).
- Coke intermixed with the calciner feed materials can serve as an effective reductant.
- FeS subsequently precipitated by reaction of Na_2S solids with dissolved $FeSO_4$ is more reactive than the thermally regenerated FeS.
- Precipitated FeS does not require milling, thereby minimizing the size of milling equipment required for the total supply of FeS to the absorber.

Thermal Regeneration

After the feed has been dried in a steam-heated co-current rotary dryer, an indirect coal-fired rotary drum kiln is used to roast and regenerate feed solids at a temperature as high as $1400^{\circ}F$. Sulfur released during the calcination leaves the kiln as a vapor and is recovered by condensation. By-product steam is generated in the sulfur condensation step, as well as in an unfired waste heat boiler fed by the calciner exhaust combustion gases.

Note that indirect calcination isolates the kiln combustion gas from the CO/CO_2 process atmosphere laden with the sulfur product vapor. Thus, unlike FGD processes such as the MgO process that use elevated temperatures requiring direct-calcination operation, the SULF-X Process by-product yield and calcined solids do not become intermixed with or contaminated by the combustion gases from calciner fuel firing.

This, in turn, affords the opportunity to use the cheapest available fuel for the calcining operation.

PROCESS CONTROL

Principal provisions for control of the process and for isolation and segregated shutdown of individual process steps are comparatively simple.

Wet Scrubbing

The scrubbing operation is controlled by maintaining a set-point pH in the scrubbing slurry through regulation of feed of the regenerated solids slurry entering the absorber loop. Additionally, feed of this slurry and bleed-off of spent slurry to the regeneration step are coordinated to maintain scrubbing slurry inventory and density in an optimum range to ensure scrubbing effectiveness. Safeguards are provided to prevent slurry-pH excursions that could impair gas cleaning efficiency or cause objectional decomposition of process chemicals.

Decoupling of Process Steps

To afford means of temporarily shutting down individual process steps for maintenance without interrupting overall system operation, a process inventory and freeboard capacity are provided for storage of process materials including:

- Dewatered spent scrubber solids in thickener-tank bottom (dryer feed)
- Dried-cake (calciner feed)
- Calciner quench-tank slurry (scrubber feed)

MATERIALS AND ENERGY UTILIZATION

A current, detailed process economics evaluation by Electric Power Research Institute (EPRI)(1) includes a preliminary conceptual design of a hypothetical SULF-X Process installation for removal of SO_2 from a 500 MW unit fired by 4.0% sulfur bituminous coal. The consumptive use of materials and energy and the yield of by-product elemental sulfur was estimated based on full-load operation at 500 MW. Assigning applicable dollar values to each item as displayed below, it is seen that the overall cost of materials and energy can be offset by by-product revenue. Thus, for one day of 500 MW operation at full load (generating 12 million KWH), the cost summary is as follows:

Consumption

Pryites, 2.8 tons/hr x 24 x $25/ton	$ 1,680
Na_2S, 0.3 tons/hr x 24 x $470/ton	3,384
Water, 727 gpm x 1440/1000 x $0.30/Mgal	314
Coke, 5.3 tons/hr x 24 x $51.30/ton	6,525
Coal, 1.3 tons/hr x 24 x $35/ton	1,092
No. 6 oil, 3 gpm x 1440 x $0.82/gal	3,542

Consumption (Continued)

H.P. steam*, 100,500 lb/hr x 24/1000 x $4/Mlb	9,648
L.P. steam, 4,500 lb/hr x 24/1000 x $3 Mlb	324
Electric power, 12,800 KW x 24 x $0.045/KWH	13,824
Total Material and Energy Cost	$40,333**

Production

Elemental sulfur, 9 tons/hr x 24 x $200/ton	$43,200
Total By-Product Revenue	$43,200**

PROCESS ECONOMICS

SO_2 REMOVAL ONLY

The detailed 1983 EPRI economic study of diverse flue gas desulfurization processes at the 500 MW scale in 4% sulfur bituminous coal service estimates the cost for the SULF-X Process as follows:(1)

SULF-X PROCESS
(in December, 1982 dollars)

Capital Cost (Including contingencies) $/KW	Levelized Busbar Cost (Capital and Operating) No By-Product Credit mills/KWH	$75/ton Sulfur Credit mills/KWH
295	22.8	20.0

FGD process selection is highly site-specific and will depend on items such as the complexities anticipated in throwaway waste management and the actual revenues available from by-product sales. The market value of the sulfur by-product significantly affects the operating cost of the SULF-X Process. For example, with the sulfur price projected at $200/ton, the levelized busbar cost with by-product credit decreases to a very favorable (within the guidelines of the study) 15.4 mills/KWH.

SIMULTANEOUS SO_2/NO_x REMOVAL

1979 in-house studies of similar utility-scale applications for commercially available combined SO_2/NO_x removal facilities achieving 80% NO_x removal confirm that SULF-X is substantially less costly than other available wet simultaneous SO_2/NO_x removal processes. Moreover, while capital cost is slightly higher than for combined removal methods that utilize dry catalytic NO_x control, the SULF-X Process is estimated to have lower annual revenue requirements than other available SO_2/NO_x removal methods including those based on dry NO_x removal.

The levelized busbar cost of a 500 MW SULF-X installation designed for simultaneous SO_2/NO_x removal achieving 80% removal of NO_x is projected to be 4 mills/KWH higher than that for SO_2-only service reviewed above. This increment, representing an increase of 20 to 25% over the cost of simple FGD capability, consists almost entirely of capital costs for added parallel-operated absorber trains required to provide increased gas residence time for high-efficiency NO_x removal.

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* 50°F indirect-type flue gas reheat

** Approximately 3.5 mills/KWH

INDUSTRIAL BOILER APPLICATIONS

To take advantage of the favorable economics of large-scale sizing, small-capacity coal-fired boilers in an industrialized region may be equipped with the wet-scrubbing-system portion of the SULF-X Process while being served as a group by a single, centrally located regeneration plant. The central plant would supply fresh calcine to each scrubber and receive the spent sulfides on the return haul. The attractiveness of the economics of such applications may approach that of utility-plant-scale installations.

SULF-X PROCESS DEMONSTRATION OPERATION

Development of the SULF-X Process through laboratory and field test work was completed in 1980. Design and construction of a 1.5 MW stoker-boiler size integrated process installation for the Commonwealth of Pennsyvlania was completed in 1982. This demonstration facility has been in round-the-clock operation since early 1983, providing data for process scale-up to larger coal-fired boiler applications.

INTRODUCTION

The integrated SULF-X Process installation completed in 1982 is now in demonstration operation at the bituminous coal-fired boiler plant of the Western Center Hospital of the Commonwealth of Pennsylvania at Canonsburg, Pennsylvania, near Pittsburgh. The system is designed for the simultaneous control of SO_2 and NO_x emissions from one of the plant boilers at a flue gas flow rate of approximately 7,000 acfm. Its current test/demonstration operation in 1983, under funding by the Commonwealth of Pennsylvania, constitutes a critical interim stage in the scale-up and application of the process through collection of steady-state operating data in anticipation of its ultimate use in large commercial installations. The continuous, integrated operation of this demonstration facility at the same time has afforded a needed opportunity to study the effects on the overall chemistry of any accumulation in the system inventory of reaction products and tramp materials, and to evaluate the compatability of the installed equipment and instrumentation.

DESCRIPTION OF WESTERN CENTER FACILITIES

The SULF-X installation at Canonsburg is a $2 million facility dedicated by the Commonwealth of Pennsylvania to a 12-month demonstration period. The system advantageously utilizes off-the-shelf process equipment and control instrumentation to provide a practical replica of the process facilities that are expected to be ultimately applied in large commercial installations. Thus, the system models a large-scale installation, using applicable mass transfer devices and mechanical equipment to duplicate primary unit operations such as the gas absorption, slurry dewatering and thermal/chemical regeneration steps, all of which are sub-processes commonly used in industry. See the Flow Diagram, Figure 3.

Flue gas cleaning equipment provided includes a 3 feet diameter radial blade booster fan, a lined carbon steel gas pre-cooler and a 73 feet high by 5 feet diameter

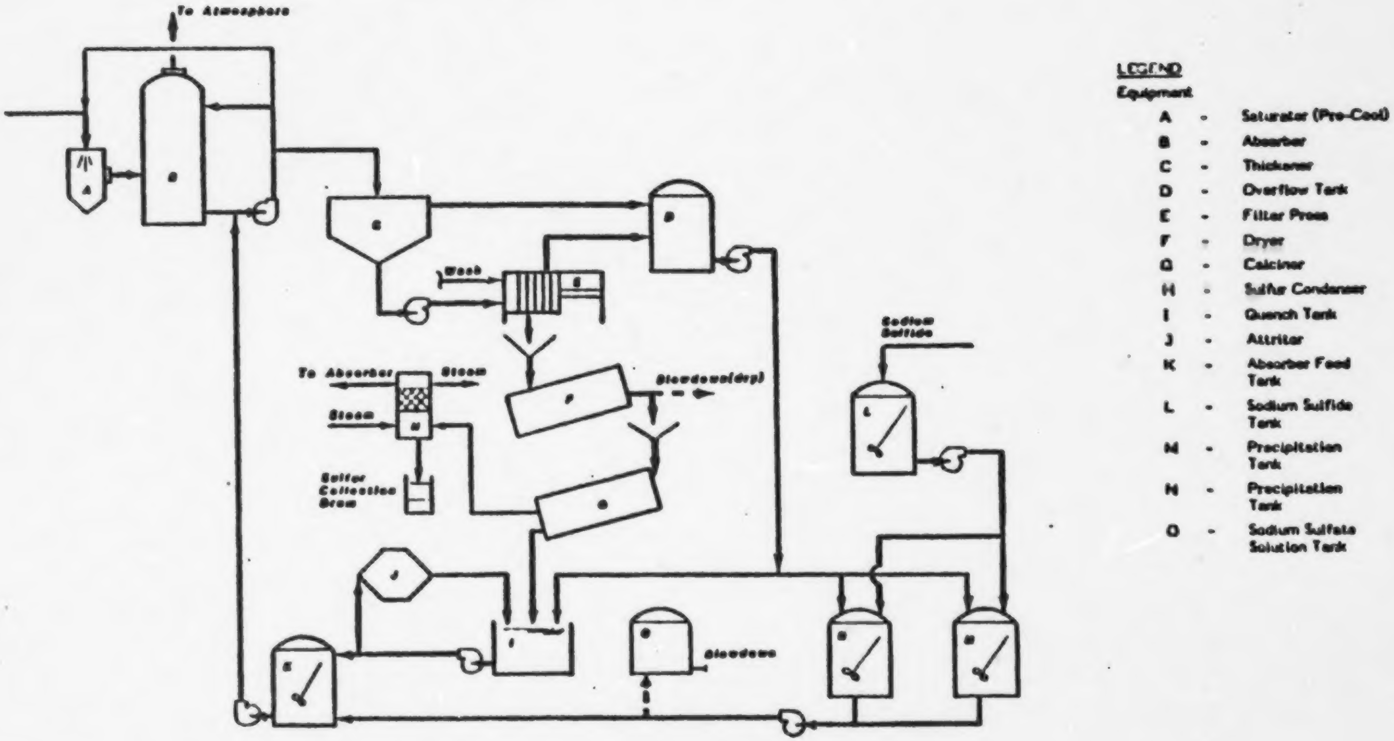

Figure 3. Flow Diagram - SULF-X System at Western Center

FRP scrubber (absorber) vessel using cross-fluted fixed plastic packing. Slurry-bleed dewatering equipment consists of a 10 feet diameter lined thickener with overflow tank, a horizontal plate-and-frame filter containing seventeen 1 meter x 1 meter vertical filter plates and a steam-heated rotary dryer. The regeneration system consists of an indirect gas-fired rotary calciner, a calcine quench tank and a steam-cooled vertical shell-and-tube sulfur condenser. For supply of regenerated reagent to the gas absorption system, the system includes an agitated-media-type calcine slurry crusher (attritor) and an absorber fresh slurry feed tank.*

Without adversely affecting the simulation of large commercial system operations, the overall demonstration program has been simplified by excluding sodium sulfate crystallization facilities. Instead, dissolved sodium sulfate is purged from the system by blowdown. Purchased sodium sulfide is used (in lieu of sodium sulfide that would otherwise be available from calcination of sodium sulfate) to sustain the quench tank reaction, which recovers dissolved ferrous sulfate by converting it to precipitated FeS reagent. To further simplify the operation of this small-capacity demonstration operation, the calciner is fired with LPG (propane) instead of coal. Since calcination is by indirect heat transfer both in the commercial scale and Western Center designs, use of alternative fuels does not affect process chemistry or the validity of the test operations. In addition, pre-cooling and water saturation of the boiler flue gas is carried out in a simple gas quenching step utilizing recirculating process slurry. In absence of an isolated pre-scrubber wastewater purge, chloride will be purged by the discard of the dissolved sodium sulfate blowdown stream. As a result of other simplifications, calciner combustion gases are exhausted without by-product steam generation and sulfur condenser off-gas is recycled to the absorber inlet instead of being treated in an incinerator.

PROBLEMS IN COMMISSIONING OF THE SYSTEM

Principal startup/operating problems encountered with the initial system installation included difficulties in mechanically transferring the filter cake to the dryer, in moving the filter cake through the dryer and in oxidation of the dried solids by air that was entering the process side of the calciner. Because of the small batch-scale nature of the filtering operation at Western Center, movement of wet filter cake is expected to continue to require special operator attention during continuous process runs. The original dryer and calciner were replaced in late 1982 with the more compatible equipment described above and these operations are now considered to be free of major problems. Adverse oxidation of wet cake was experienced in early operation of the new steam-heated rotary dryer, but this problem has been eliminated. Retrofit modifications have also been made to the sulfur vapor vacuum and condensing systems to avoid solidification of sulfur before and after the condenser.

OVERVIEW OF SYSTEM TESTING WORK

The Western Center absorber has successfully demonstrated up to 99+% removal of SO_2 and up to 93% removal of NO_x in 2%-sulfur bituminous coal operation. At a superficial velocity no greater than 2 feet per second, NO_x removal efficiency of 90% is obtained with essentially 100% SO_2 removal. Figure 4 shows the graphical relationship

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*For initial ferrous chemical charge and subsequent Fe makeup supply, the Western Center operation has used copperas produced from waste pickling liquor neutralization.

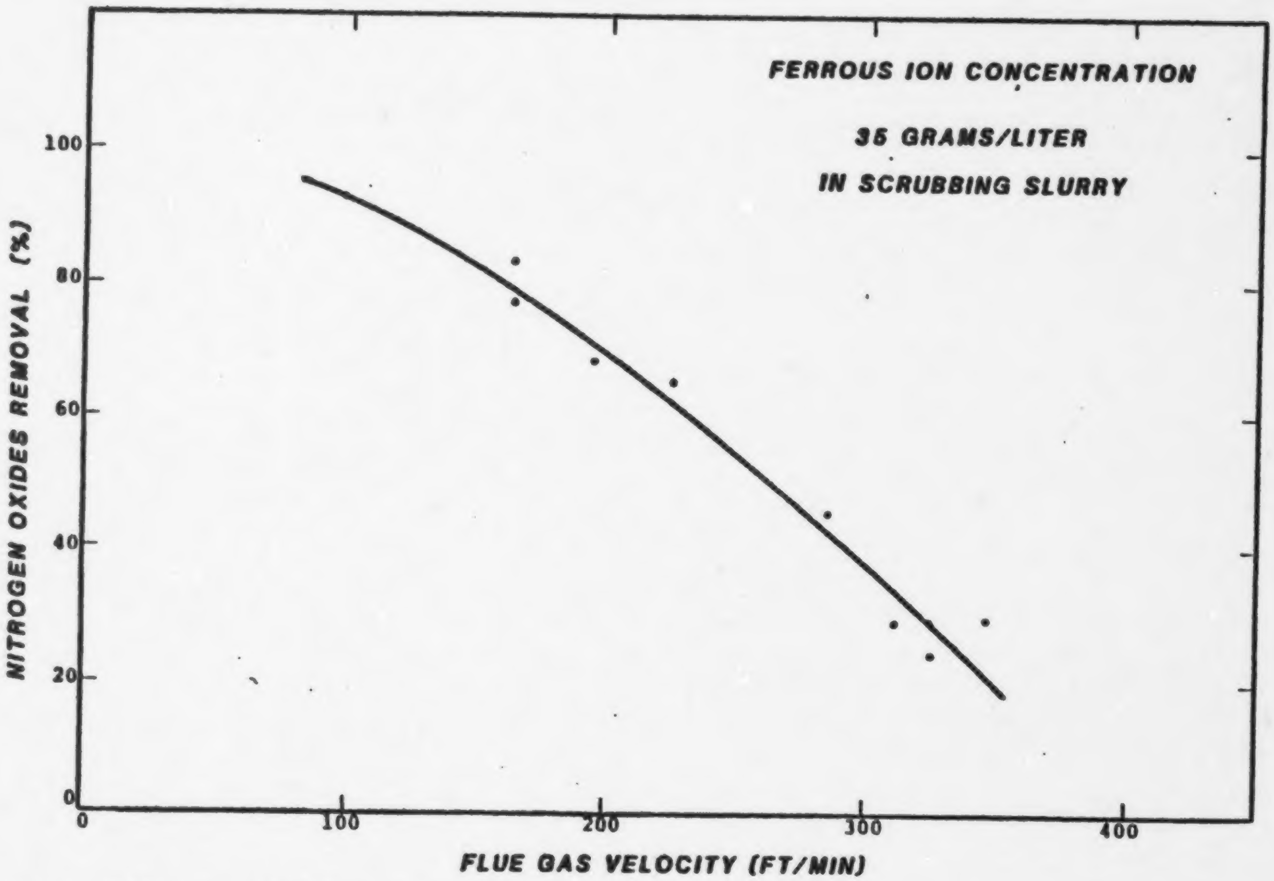

Figure 4. Nitrogen Oxides Removal vs. Superficial Flue Gas Velocity

between NO_x removal and absorber superficial gas velocity for the present absorber configuration functioning as a simple single-stage counter-current contactor. Other variables affecting NO_x removal rates are slurry temperature and viscosity and FeS particle size. Figure 5 shows the inlet SO_2 flue gas concentrations and the percent absorbed when the system was operated for simultaneous SO_2 and NO_x removal. At a superficial gas velocity in the scrubber of approximately 10 feet per second and a liquid/gas ratio of 60 U.S. gallons per thousand cubic feet, typical of operation of commercial FGD systems, SO_2 removal at an efficiency of 95+% is accompanied by only token NO_x removal.

The absorption system has operated automatically with the pH controller regulating fresh slurry addition and spent slurry removal to maintain a constant, pre-set pH. Spent slurry has dewatered well in the thickener and filter, and the FeS precipitation and chemical makeup systems have performed as designed. Comprehensive data are being collected for a broad range of operating conditions to achieve parametric testing that will optimize system design for NO_x removal levels up to 80%, and to yield a data base to be used for scale-up.

CONTINUOUS PROCESS OPERATIONS

Chemistry of Steady-State Process Streams

The iron, sulfur and sodium species in the recycled scrubbing slurry were monitored throughout the project for both daily process control and overall system performance evaluation. Relatively high concentrations of sodium sulfate (Na_2SO_4) were measured but were anticipated since the sulfate crystallizer had not been included as part of this system and a low liquid purge rate had been maintained. This rate of liquid waste blowdown was adjusted to maintain the dissolved Na_2SO_4 concentration in the slurry below its saturation point.

Analyses indicated that steady-state conditions had been achieved by the fourth operating month. Chemical consumptions and product concentrations have been consistent with earlier studies and the previously identified chemical reactions. Approximately 17% more ferrous sulfide was reacted in the absorber than was expected. This increase is believed to have been caused by the unusually high oxygen content in the flue gas, ranging from 10 to 14%. A typical analysis of process slurry is presented in Table 1.

Quantification of Solid and Liquid Waste Discharge

There was no significant solid waste discharged from the system. Approximately 13,600 gallons of sodium sulfate liquid waste were purged from the system. Analysis of this waste liquor is as follows:

Iron, Fe^{++}	0.6 g/l	Sulfate ($SO_4^=$)	76.9 g/l
Sodium, Na^+	34.0	Sulfite ($SO_3^=$)	0.1
Other cations (Ca, Mg, Al)	7.6	Thiosulfate ($S_2O_3^=$)	9.3
. . . . . .		Chloride (CL^-)	1.0
pH	6.4		

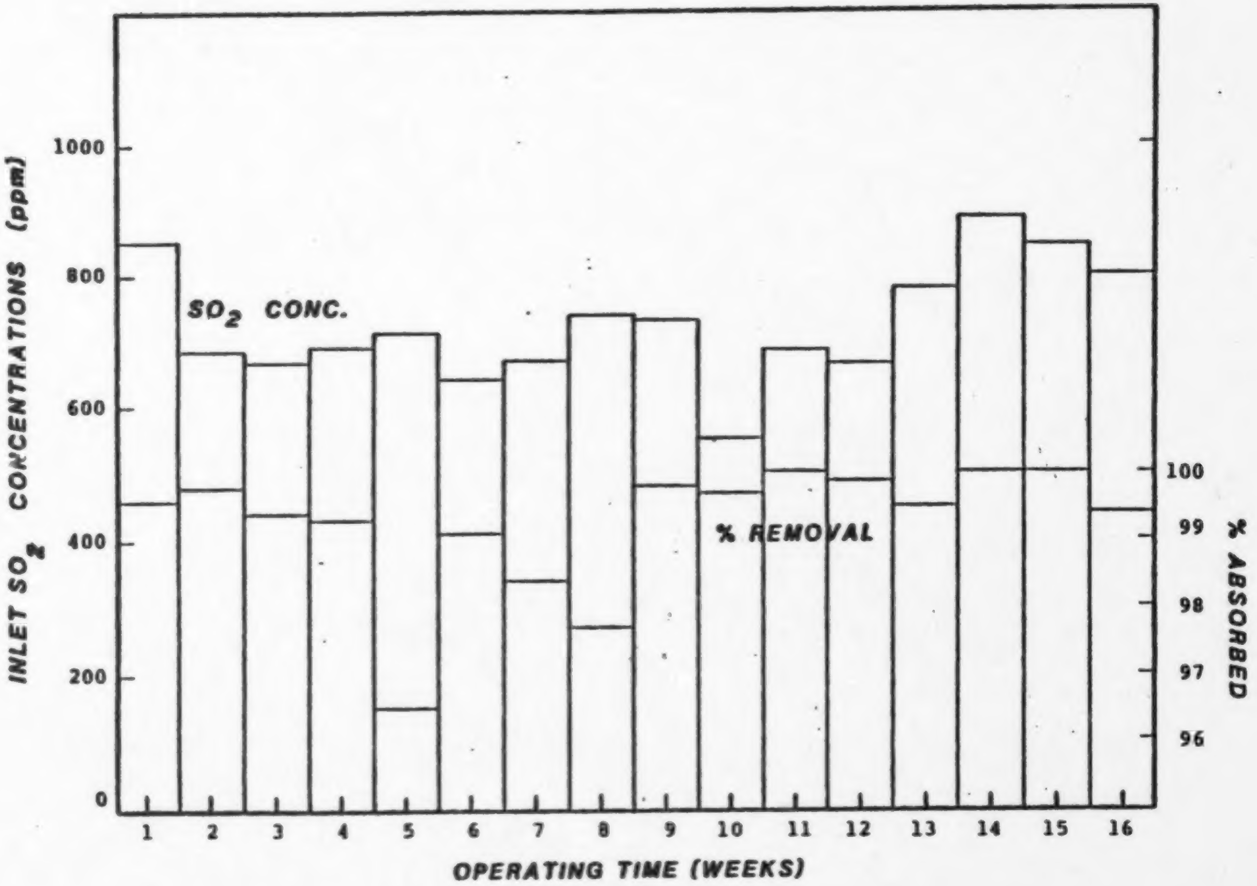

Figure 5. Inlet Flue Gas SO_2 Concentrations and Percent Absorbed by SULF-X System

TABLE 1. SULF-X PROCESS SIMULTANEOUS SO_2/NO_x REMOVAL

Recycled Scrubber Slurry Composition	
Iron, total	6.42%
Sulfur, total	7.25%
% Suspended solids	5.72%
% Dissolved solids	22.1 %
Specific gravity, slurry	1.29
pH, slurry	5.4

Dissolved Solids, grams/liter

Iron (Fe^{++})	49.9 g/l	Sulfate ($SO_4^{=}$)	164.0 g/l
Sodium (Na^{+})	48.0	Sulfite ($SO_3^{=}$)	0.1
Other Cations* (Ca, Mg, Al)	0.3	Thiosulfate ($S_2O_3^{=}$)	20.7
		Chloride (Cl^{-})	1.7

* Calcium and magnesium salts are contained as anti-caking agents in the purchased iron sulfate makeup; aluminum is leached from the scrubber fly ash accumulating in the slurry.

Taking into account the change in system chemical inventory, the quantity of the dissolved sulfur species in excess of that contributed by ferrous sulfate makeup represents approximately 7% of the sulfur dioxide absorbed during the test period. Sulfur (sulfate) losses in the blowdown can be expected to significantly decrease and the ratio fo chloride to dissolved sulfur species correspondingly increase for a SULF-X system installation that includes a sodium sulfate crystallizer.

SUMMARY OF FINDINGS IN WESTERN CENTER DEMONSTRATION RUNS TO DATE

The installation and operation of the demonstration SULF-X facility for the Commonwealth of Pennsylvania has been an important and successful step in the continuing development of the Process. The data and information collected during all phases of the project are valuable additions to the engineering and operating criteria needed for building a larger SULF-X FGD installation of optimized design.

During the operation, equipment and instrumentation compatability in the face of variances in process operation was thoroughly evaluated. The results served to identify means of overcoming potential problem areas and confirmed the accuracy of previously established process design criteria:

- The packed tower functioned as an effective mass transfer unit and should be used in future systems requiring substantial NO_x removal.
- Use of pH for control of process chemistry and slurry make-up rate was effective and reliable.
- Dithionates and imidodisulfonates did not accumulate in the recirculated slurry nor in the sodium sulfate solution waste.
- Slurry bleed from the absorber was readily dewatered with a pressure filter after thickening. This final dewatering could probably be performed with a centrifuge.
- Material transfer of the filter cake was troublesome and requires additional investigation if batch-wise filtration is to be used reliably in routine system operation.
- To ensure efficient regeneration and sulfur recovery, precautions must be taken to avoid excessive oxidation of the solids during the drying and calcining steps.
- Iron oxide was converted in the calciner to an active ferrous form, thereby substantially decreasing the anticipated waste solids blowdown rate.

PROJECTED PROCESS DEMONSTRATION AT LARGER SCALE

A new and larger scale demonstration of the process is now planned in conjunction with future development of process licensees for commercial application of SULF-X technology in the U.S.A., Canada, Japan, West Germany and other industrialized countries. In a new system installation for this purpose, it is anticipated that a pulverized coal-fired boiler facility at the 10 to 60 MW size range would provide a minimum-risk basis for ultimate scale-up of the process to utility plant applications larger than 300 MW. At the same time, it should be possible to reliably quantify the benefits, simplifications and anticipated economies in the process design resulting from typically low flue gas oxygen content in pulverized-fired service.

EXPANDED WESTERN CENTER DEMONSTRATION FOR U.S. DEPARTMENT OF ENERGY

A new test/demonstration program is now being carried out at Western Center under the sponsorship of DOE to define optimum design/operating conditions for achievement of 90% NO_x removal in simultaneous SO_2/NO_x scrubbing operation. A systematic sequence of test runs is being conducted in a continuous integrated closed-loop mode to evaluate performance, identify important operating variables and provide data for assessing the technical and economic feasibility of high efficiency, simultaneous SO_2/NO_x removal. Additionally, process runs are being made to demonstrate regeneration of sodium sulfate with coke as a reductant. Alternative fuel and reductant materials are also being investigated.

90% NO_x REMOVAL

The piping of the two in-series absorber beds, originally arranged to operate as a simple single-stage counter-current scrubber, is being modified to provide two separately operated scrubbing stages. A series of test runs to evaluate performance and study important operating variables will be conducted. Parameters to be adjusted and assessed include liquid/gas flow ratio of the individual absorber stages, superficial gas velocity and chemistry of inlet slurry to each stage. Parametric testing will take into account the effect of other variables that may significantly affect the NO_x absorption rate.

RUNS WITH MODIFIED CALCINER OPERATION

The calciner test program will model-test typical large-scale SULF-X Process operation in which the regeneration system is regenerating all tramp sulfate formed in the scrubbing step to fresh, usable sulfide reagent. This will entail continuous addition of coke reductant and sodium sulfate solids at the calciner feed inlet. Individual test runs will be made to investigate alternative use of petroleum coke, metallurgical coke and boiler-house coal as the reductant.

LONG-TERM RELIABILITY RUN

After establishing an applicable mode and optimum conditions for SULF-X operation at 90% NO_x removal, a continuous 30-day system reliability run will be made to verify the design basis for such commercial operation. Quantities and quality of waste discharges during sustained operation will be monitored to assess system waste production characteristics and to identify any specific provisions that would be needed for treatment and ultimate disposal of the waste.

CONCEPTUAL COMMERCIAL-SCALE DESIGN FOR 90+% SO_2, 90% NO_x REMOVAL

Based on data generated and conclusions drawn during the field test work, a conceptual SULF-X Process design will be made for a hypothetical 500 MW power plant burning high sulfur bituminous coal. This design will serve as a basis for assessment of the technical and economic feasibility of this mode of operation of the SULF-X Process.

CONCLUSIONS

SULF-X is an effective new wet-scrubbing-type regenerative flue gas desulfurization process applicable to all fossil fuels. It has been demonstrated to advantageously use and regenerate ferrous sulfide originating from industrial iron by-product wastes to convert flue gas SO_2 into usable elemental sulfur while producing negligible solid and liquid waste except common chloride liquor purge.

The process has demonstrated an inherent versatility in functioning as either a system for SO_2 removal only, or for simultaneous SO_2/NO_x treatment. Cost projections indicate that it is an economically attractive alternative to throwaway-waste-type FGD system where a regional market for by-product sulfur exists, particularly in applications where solid waste management costs would be expected to be disproportionately high. Moreover, the capital investment for small to medium-sized SULF-X units can be minimized if concentration of industry in a region allows a central regenerating plant to serve a number of individual gas scrubbing units.

Additionally, the SULF-X system is technically and economically attractive where nitrogen oxides removal is also necessary at present or in the future. The system lends itself to being installed and operated as an SO_2-only FGD system and then later being readily modified to also remove nitrogen oxides.

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